RECLAMATION

Managing Water in the West

Desalination and Water Purification Research and Development
Report No. 97

Comparison of Advanced Treatment Methods for Partial Desalting of Tertiary Effluents

Samer Adham, Ph.D., Thomas Gillogly, Ph.D., Geno Lehman
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Agreement No. 99-FC-81-0189

U.S. Department of the Interior
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U.S. Department of the Interior
Mission Statement

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GLOSSARY

Abs	absorbance
ASME	American Society of Mechanical Engineers
BAT	best available technology
Bureau	U.S. Bureau of Reclamation
°C	degrees celsius
cfu	colony forming unit
CIP	clean-in-place
CMF-S	continuous microfiltration – submerged
District	Santa Clara Valley Water District
EDR	electrodialysis reversal
FWR	feed water recovery
GAC	granular activated carbon
gfd	gallons per square foot per day
MF	microfiltration
MMF	multimedia sand filtration
MPN	most probable number
NDP	net driving pressure
NTU	nephelometric turbidity unit
PLC	programmable logic controller
PP	polypropylene
psi	pounds per square inch
PVdF	polyvinylidene fluoride
QA	quality assurance
QC	quality control
RO	reverse osmosis
SBS	sodium bisulfite
SBWR	South Bay Water Recycling
SDI	silt density index
TDS	total dissolved solids
TMP	transmembrane pressure
TPS	Transmission Pumping Station
TSS	total suspended solids
UF	ultrafiltration
USEPA	United States Environmental Protection Agency
SJ/SC WPCP	San Jose/Santa Clara Water Pollution Control Plant

1. EXECUTIVE SUMMARY

South Bay Water Recycling (SBWR) is a program of the San Jose/Santa Clara Water Pollution Control Plant (SJ/SC WPCP) that supplies recycled water for irrigation and industrial use the Silicon Valley area of Northern California. Beginning full-scale operation in 1998, SBWR supplies over 10 mgd of recycled water to more than 400 customers. The concentration of total dissolved solids (TDS) ranges between 770 and 820 mg/L such that the water is suitable for all current uses. However, as recycled water is used more extensively for evaporative cooling the TDS will likely increase to the point that the water may no longer be suitable for irrigation of some salt-sensitive plants, and the cost of pretreatment for industrial use will increase.

This study investigated the feasibility of using two advanced water treatment alternatives – microfiltration followed by reverse osmosis (MF/RO) and electrodialysis (EDR) – to reduce the salinity of recycled water from the SJ/SC WPCP from a concentration of 750±50 mg/L TDS to 500 mg/L (38 percent reduction) and 350 mg/L (56 percent reduction). Pilot scale equipment for the two treatment alternatives was provided by two separate vendors and operated for approximately six months.

<u>Feasibility of Microfiltration/Reverse Osmosis</u>
Membrane pretreated water was continuously fed to the RO pilot system at a flow rate of approximately 20 gpm. MF/RO Pilot testing included two phases of operation to evaluate RO performance at an applied flux of 15 gfd and feed water recovery of 50-65 percent. During Phase I, polypropylene (PP) MF membranes were used for pretreatment of tertiary effluents before being fed to the RO pilot. At a flux of 16 gfd, the PP membranes were effective at reducing the silt density index (SDI) from 8 to less than 1 during operation (>1000 hours) such that the RO membranes produced water with a TDS below 10 mg/L (98 percent salt rejection).

Testing of the MF/RO using polypropylene membranes for pretreatment was terminated due to MF membrane damage by free chlorine in the feed. As a corrective measure, PP membranes were replaced with new chlorine-tolerant polyvinylidene fluoride (PVdF) membranes for pretreatment before RO. At an increased flux of 30 gfd, the PVdF membranes operated for an additional 500 hours and maintained the SDI below 1.0 with no operational failures were observed. Similarly, up to 97 percent salt rejection was observed with the RO pilot during Phase II. Results from Phase I and II confirmed that MF/RO is a suitable advanced treatment technology for the reclaimed water produced by the SJ/SC WPCP.

<u>Feasibility of Electrodialysis Reversal (EDR)</u>
The two primary objectives of the EDR pilot testing were to evaluate system performance of the EDR to produce 350 and 500 mg/L TDS water, as well as assess the impact of three different alternative pretreatment methods [Granular Activated Carbon (GAC)/Multimedia Filtration (MMF), Microfiltration (MF) and, Cartridge Filtration only]. Pilot testing results demonstrated that the EDR was capable of meeting the desired treatment goals for all pretreatment methods tested. Because of the high quality effluent produced by the SJ/SC WPCP (i.e., low suspended solids, turbidity <1.0 NTU, low plant chemical residuals), the EDR performed well without pretreatment (cartridge filtration only). However, it is important to note that these conclusions were based short-term testing results (< 500 hours per pretreatment condition), primarily focused

on comparing the three evaluated pretreatment methods to EDR membranes. Long term testing would be required to fully investigate the O&M impacts of operating EDR membranes without significant pretreatment.

Relative Costs

The cost of producing 50 MGD of partially desalinate recycled water from the San Jose/Santa Clara Water Pollution Control Plant through MF/RO and EDR treatment is shown in the table below. Treatment costs expressed per thousand gallons ($/kgal) include capital and operating expenses calculated for production of a blended product water with a final salinity of either 350 mg/l or 500 mg/L.

	Cost of Treatment (including Capital and O&M), $/kgal	
Type of Treatment	*350 mg/L*	*500 mg/L*
MF/RO	$0.86	$0.51
MF/EDR	$0.85	$0.55
*EDR**	$0.57	$0.32

* EDR with cartridge filtration only

Recommendations for Future Study

This study has proved useful to the sponsoring local agencies in two ways. In the first place, the data obtained in this investigation will support future decisions on advanced water treatment. For example, since the study indicates that electrodialysis is the lower cost alternative, the agencies could facilitate the design of full-scale facilities by validating the pretreatment requirements and cost estimates reported here. On the other hand, if reverse osmosis is preferred as a means of removing nonionic contaminants (e.g. pharmaceutically active and endocrine disrupting compounds) this study will allow the agencies to estimate the additional cost required to remove these emerging pathogens over and above the cost to reduce salinity. Either way, the resultant data is a useful step towards selecting the most appropriate technology for improving the quality of recycled water in this area.

Second, the study has provided valuable experience for local treatment plant operators and managers in the reliable operation of advanced treatment equipment. The variety of operational challenges that arose during this brief investigation gave plant personnel exposure to a wide range of issues, including prevention of fouling, chemical pretreatment, electrical safety and telemetry all of which have been documented in this report. As a result, the agencies are better able to evaluate alternative advanced water treatment designs and better prepared to operate and maintain future facilities. The City of San Jose is particularly grateful to the Bureau of Reclamation for their early and continued support of this work, as well as to the other project co-sponsors listed in Section 2.2.

Based on the results of this investigation, additional pilot testing is recommended to confirm long-term performance of EDR membranes without pretreatment as well as for increased throughput of MF/RO under optimized conditions for the relatively high-quality reclaimed water produced by the SJ/SC WPCP. Such an evaluation should be designed to provide data on the following parameters:

- operational experience including maintenance requirements
- evaluation of membrane performance during plant “upsets”
- full-scale design criteria including a refined cost analysis
- operator training

2. INTRODUCTION

2.1 Background

South Bay Water Recycling (SBWR) is a program of the San Jose/Santa Clara Water Pollution Control Plant (SJ/SC WPCP) that supplies recycled water for landscape irrigation and industrial use to three cities in the Silicon Valley area of northern California. This study was designed to evaluate the feasibility of improving recycled water quality through the use of two advanced treatment technologies, microfiltration followed by reverse osmosis (MF/RO) and electrodialysis reversal (EDR).

The SBWR system was constructed primarily to reduce effluent discharges into the south end of San Francisco Bay. In 1989, the San Francisco Regional Water Quality Control Board limited the plant's discharge to 120 mgd when they determined that plant effluent converted salt marsh to fresh marsh and reduced the habitat of two endangered species—the salt marsh harvest mouse and the California clapper rail. In response to this order, the cities of San Jose and Santa Clara (joint owners of the plant that also serves six other cities and three sanitary districts) prepared the South Bay Action Plan. The plan consisted of three components—1) water conservation, 2) marsh mitigation and 3) water reuse. The water reuse component was accomplished through the construction of a 60-mile recycled water distribution system including four pump stations and a reservoir, with a capacity to distribute peak recycled water flows of up to 50 mgd. The project was partially funded by a construction grant from the U.S. Bureau of Reclamation (Bureau) through its Title XVI program (PL102-575).

SBWR began full-scale operation in 1998 and now supplies over 10 mgd of recycled water during the summer months to more than 400 customers for landscape irrigation and industrial use. Concentration of total dissolved solids (TDS) ranges between 770 and 820 mg/L such that the water is suitable for all current uses. However, the TDS will likely increase to the point that recycled water may no longer be suitable for irrigation of some salt-sensitive plants, and the cost of pretreatment for industrial use increases as it is used more extensively for evaporative cooling.

2.2 Objectives of the Study

In 1999 the Bureau awarded the City of San Jose a research grant (BOR #99-FC-81-0189) to investigate the feasibility of using advanced water treatment to reduce the salinity of recycled water for industrial use. Montgomery Watson Harza (MWH) was selected a principal investigator for the study, and additional funding was obtained from the Santa Clara Valley Water District (in conjunction with the Metropolitan Water Districts of Southern California) and the WateReuse Foundation. USFilter and Ionics Ultrapure Water Corporation (Ionics) provided pilot water treatment equipment that was used in this investigation.

The purpose of this pilot study was to compare non-thermal demineralization processes for the partial desalination of non-potable recycled water and determine if excessive membrane fouling, process interruption/failure, etc. existed with the water quality of tertiary treated wastewater produced by the SJ/SC WPCP. MF/RO and EDR were selected to reduce effluent salinity from a concentration of approximately 750±50 mg/L TDS to 500 mg/L (38 percent reduction) and 350 mg/L (56 percent reduction). The study optimized the operating parameters for each process (RO and EDR) with respect to efficiency, reliability, effectiveness and cost.

2.3 Advanced Water Treatment Processes

These two non-thermal demineralization processes were chosen as they are recognized as viable alternatives for demineralization of tertiary treated wastewater to produce water that could be used for non-potable applications, including industrial and landscape irrigation uses. With further treatment, such partially desalinated water could also be adapted to specialized applications (like ultra-pure water for manufacturing electronic products). The different treatment processes were analyzed to determine the most cost-effective demineralization technology.

2.3.1 Reverse Osmosis

Reverse osmosis is a pressure driven membrane separation process where dissolved solutes are separated from the solution by forcing the water through a semi-permeable membrane under a pressure higher than the osmotic pressure of the solution. The most common type of reverse osmosis membrane module used is the spiral-wound configuration. As shown in Figure 2-1, two sheets of the membrane are placed back to back, separated by a spacing fabric that acts as a permeate channel. Three sides of the sheet are glued together to form the envelope or leaf. The open end of the leaf is attached to the central permeate tube. A feed stream spacer is placed between a pair of membrane leafs to allow the feed water to flow across the membrane surface. Finally, the leafs and feed spacers are spirally rolled into a cylindrical shape and sealed to create a tightly wound element.

Individual reverse osmosis membrane elements are housed in cylindrical pressure vessels. As shown in Figure 2-2, feed and concentrate flow through the feed-side channels in a straight path parallel to the direction of the permeate collection tube. Water penetrates the membrane and is collected in the center permeate tube. The remaining water passes the element and exits through the concentrate port of the pressure vessel. Typically, several elements are housed in series in a pressure vessel in which the concentrate from one element serves as the feed to the next in series.

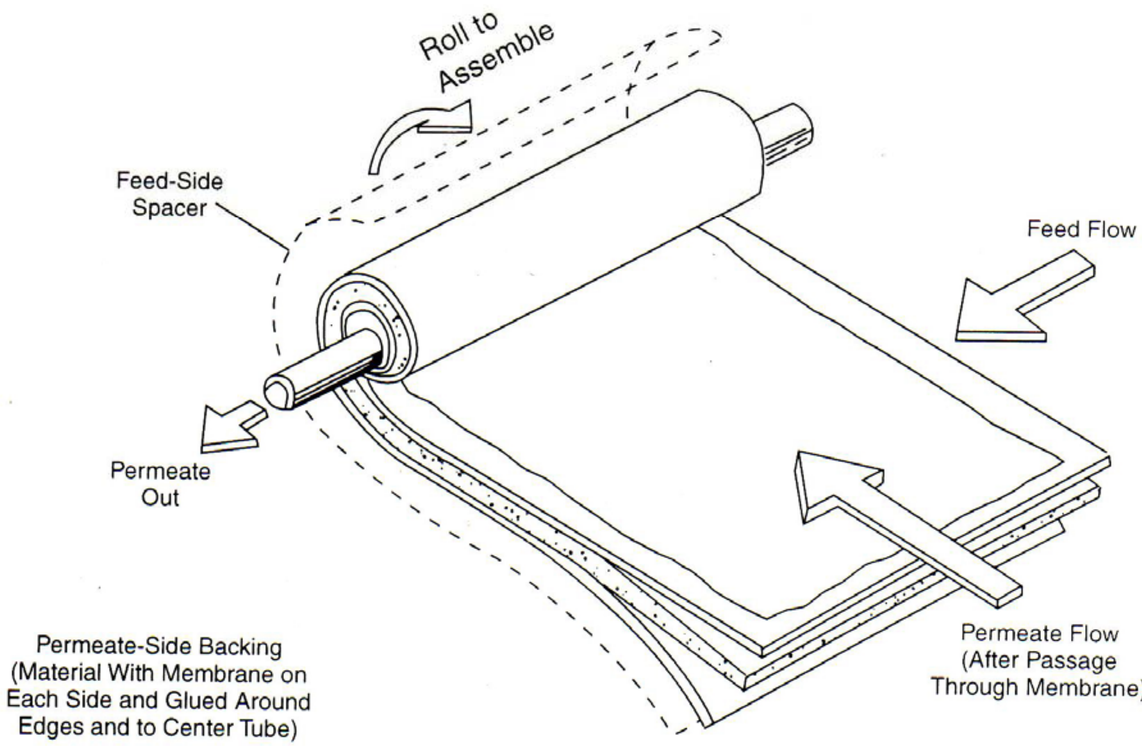

Figure 2-1. Reverse osmosis spiral wound module (AWWA 1999)

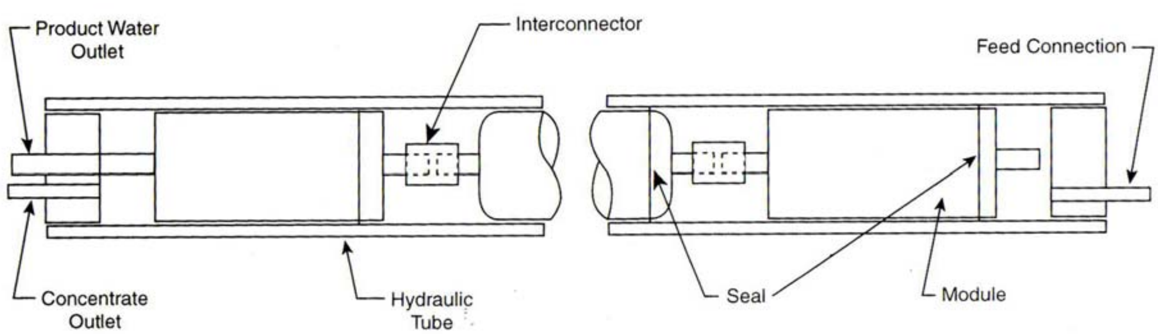

Figure 2-2. Reverse osmosis pressure vessel assembly (AWWA 1999)

Reverse osmosis has been selected as a best available technology (BAT) by the USEPA for the removal of inorganics such as sulfate and nitrate that can comprise a large percentage of the TDS present in a water or wastewater. It has been extensively tested for treatment of reclaimed water and several full-scale facilities have been constructed. A list of recent literature references discussing RO full-scale installations, operational experience, and applications are provided in Appendix B.

2.3.2 *Electrodialysis Reversal*

In the electrodialysis removal process, charged ions are removed from solution by applying an electrical potential across a stream of water. This causes the ions to move towards the opposite charged electrode (Figure 2-3). Ion selective membranes separate the stream from the electrode allowing only positive or negatively charged ions to pass through. These membranes are arranged alternately, with an anion selective membrane followed by a cation selective membrane. A spacer sheet is then placed between these two membranes forming channels in the EDR cell. As the electrodes are charged and feed water flows along the product water spacer at right angles to the electrodes, the anions (like chloride and carbonate) in the water are attracted and diverted through the anion selective membrane towards the positive electrode. This dilutes the salt content of the water in the product water channel. The anions pass through the anion selective membranes but cannot pass through the cation selective membrane and hence the anions are concentrated in the brine channel. Similarly, cations like calcium and sodium under the influence of the negatively charged electrode, pass through the cation selective membrane and are trapped in the brine channel on the other side. This results in concentrated and dilute solutions being created in the spaces between the alternating membranes.

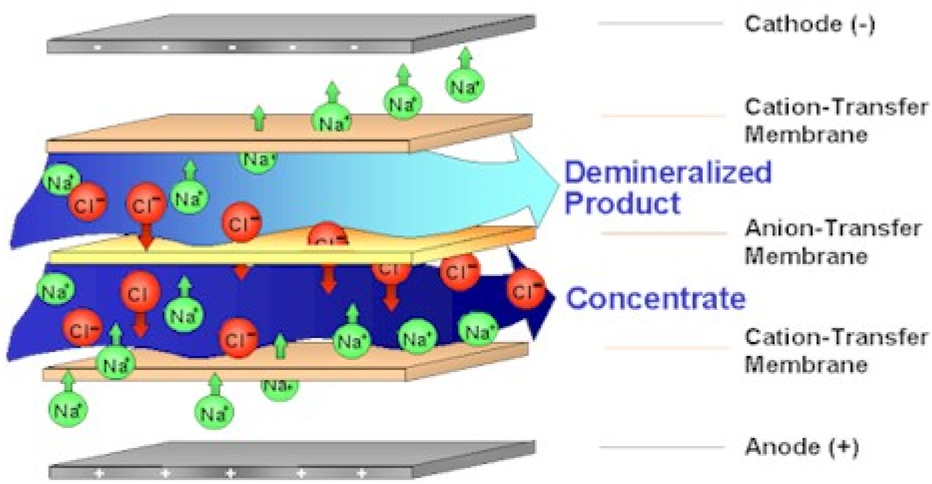

Figure 2-3. Schematic to illustrate the electrodialysis (ED) process (Ionics Inc.)

These spaces, bound by two membranes (one cationic and one anionic) are called cells. The cell pair consists of two cells, one from which the ions migrated (dilute cell for product water) and the other in which the ions concentrate (the concentrate cell for the brine). The basic EDR unit consists of several hundred cell pairs bound together with electrodes on the outside and is referred to as a membrane stack. Feed water passes through the feed paths in parallel providing a continuous flow of desalted water and concentrate from the stack.

Currently, the City of San Diego's North City Water Reclamation Plant (NCWRP) operates a full-scale demineralization facility utilizing EDR technology to reduce the salinity of reclaimed water. The water quality after tertiary treatment is similar to that of the SJ/SC WPCP and minimal pretreatment (only cartridge filtration) is being used. During the first year of full-scale operation, extensive EDR membrane fouling occurred and frequent CIP (clean in place) cleanings (even EDR membrane replacements) were required to maintain membrane integrity.

The NCWRP and Ionics determined that the EDR failures were associated with excess amounts of alum that was fed to the secondary and tertiary processes to help reduce the TSS. The excess coagulant ended up getting into the membrane stack, causing them to become severely damaged. Additionally, it was later discovered that a nearby agency was routinely dumping clarified sludge down the sewer directly adjacent to the wastewater plant. As a result, this dumping caused a plant upset condition and excessive chemical feeds were required to treat the wastewater. Consequently, when this plant upset condition occurred, the EDR membranes were damaged shortly afterwards and needed to be replaced.

Today, the NCWRP uses ferric chloride, instead of alum. Operation of the EDRs has also been modified such that if an "upset" condition is experienced (i.e., NTU>2, excess chemicals feed, etc), the EDR is temporarily bypassed for that "upset". After implementing the aformentioned process changes, EDR operation has been successful for 3.5 years, and a normal (less aggressive and less costly) maintenance schedule has been followed (Chou 2004, Reahl 2004). In addition to this recent experience with EDR, a list of literature references discussing EDR full-scale installations, operational experience, and applications are provided in Appendix B.

2.3.3 Pretreatment

An important aspect of the advanced treatment of recycled water is the selection of the pretreatment process. Membrane filtration pretreatment has been found to be the ideal pretreatment for selected RO processes in studies conducted by MWH at San Diego (MWH 1997, DRIP 2002). Pretreatment of the recycled water using microfiltration or ultrafiltration helps in particle removal and provides a higher quality of feed water to the advanced treatment process as compared to conventional pretreatment processes. Additionally, since reclaimed water quality can be highly variable, membrane pretreatment processes provide additional benefits as the product water quality from the membranes are not dependent on the feed water quality.

3. Conclusions and Recommendations

3.1 Operational Performance

3.1.1 Reverse Osmosis

- The RO was operated for more than 2000 hours at an applied flux of 15 gfd with a FWR of 50 to 65 percent using 12 RO membrane elements (DOW BW30-4040) configured in a 3 vessel, 2:1 array.
- The RO membranes tested achieved excellent salt rejection (> 98 percent).
- Membrane pretreatment (for the RO) using microfiltration membranes reduced influent turbidity from 1 NTU to 0.1 NTU.
- Two types of microfiltration membranes were evaluated: polypropylene and polyvinylidene fluoride. Both met treatment goals. However, PVdF membranes performed due to chlorine-tolerant characteristic.
- MF fluxes up to 30 gfd were used without excessive membrane fouling
- Silt density index was consistently reduced from up to SDI 20 to below SDI 1 by the MF pretreatment throughout the course of testing.

3.1.2 Electrodialysis Reversal

- The Ionics EDR pilot system operated for more than 2500 hours as a single stack configuration with up to two electrical stages.
- The EDR system achieved salt rejections ranging from 26 to 57 percent
- Stable operation was achieved during operation of the EDR without significant EDR membrane fouling or operational errors.
- Three different pretreatment scenarios were evaluated for the EDR pilot, including granular activated carbon followed by multimedia sand filtration (GAC/MMF), membrane microfiltration (MF), and cartridge filtration only.
- It was found that the EDR can be operated using a minimal amount of pretreatment (cartridge filtration only) without excessive fouling, operational failure, or a decrease in product water quality.

3.2 Costing Analysis

- Cost estimates ($/1000 gal) for 50-MGD MF/RO advanced treatment were estimated for a blended product of 350 and 500 mg/L, below.
- As GAC for chlorine reduction is considerably more expensive than chemical (i.e., SBS) reduction), a separate cost for GAC pretreatment to EDR was not considered.

	Cost of Treatment (including Capital and O&M), $/kgal	
Type of Treatment	*350 mg/L*	*500 mg/L*
MF/RO	$0.86	$0.51
MF/EDR	$0.85	$0.55
*EDR**	$0.57	$0.32

* EDR with cartridge filtration only

3.3 Recommended Future Work

Building on knowledge from other projects involving the partial desalting of tertiary effluents, this study has demonstrated that both reverse osmosis and electrodialysis reversal systems can successfully treat tertiary effluent containing high TDS salts. It was further determined that the EDR system can treat the high quality recycled water produced by the SJ/SC WPCP with only cartridge filtration pretreatment such that the EDR would be more cost-effective than MF/RO for producing a comparable volume and quality of partially desalted water. This conclusion is based on the successful operation of the EDR unit for about three weeks, during which time it is believed that a tendency towards fouling in the future would have been evident. However, it is further recommended that long-term performance of EDR membranes with and without pretreatment and of MF/RO under optimized condition occur to provide the following information:

- long-term operation data
- operation and maintenance requirement
- full-scale design criteria
- evaluation of membrane performance during planned and unplanned plant “upsets”
- refined cost analysis
- operator training

4. Materials and Methods

This section contains information concerning the materials and methods used in performing this study. The results of the study are reported and discussed in a subsequent section (Section 5, Results and Discussion), while details concerning the operation of the pilot equipment are included in the appendix (Appendix A, San Jose Operator Experience During Pilot Study).

4.1 Testing Site

The test site was located at the City of San Jose/Santa Clara Water Pollution Control Plant (SJ/SC WPCP) – South Bay Water Recycling (SBWR) Transmission Pumping Station (TPS) at 700 Los Esteros Road in San Jose, California.

4.1.1 Site Background Information

The SJ/SC WPCP is one of the largest advanced wastewater treatment facilities in California. This facility treats and cleans the wastewater of over 1.5 million people that live and work in the 300-square mile area encompassing San Jose, Santa Clara, Milpitas, Campbell, Cupertino, Los Gatos, Saratoga, and Monte Sereno.

The SJ/SC Water Pollution Control Plant has the capacity to treat 167 million gallons of wastewater per day. It is located in the Alviso neighborhood of north San Jose, at the southernmost tip of the San Francisco Bay. The SJ/SC WPCP treatment train includes the following processes:

1. Pretreatment (screening, sedimentation and grit removal)
2. Primary settling and scum removal
3. Flow equalization
4. Secondary biological nutrient removal consisting of a four-chamber aerobic/anoxic suspended growth activated sludge treatment with partial denitrification
5. Gravity filtration with anthracite coal and sand
6. Chlorine disinfection followed by sulfur dioxide dechlorination (prior to discharge) or gaseous chlorine rechlorination (prior to reuse).

The Plant is designed to remove more than 98% of biochemical oxygen demand (BOD) and more than 99% of Total Suspended Solids (TSS). The current advanced wastewater treatment plant has the capacity to treat up to 167 mgd. The flow diagram is presented in Figure 4-1.

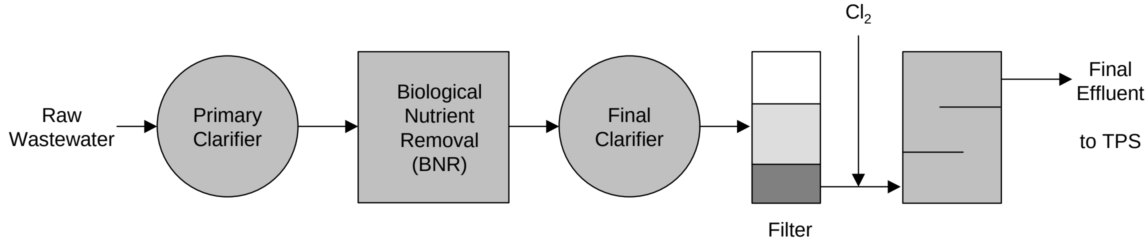

Figure 4-1. San Jose / Santa Clara Water Pollution Control Plant advanced wastewater treatment flow diagram

Most of the final treated water from the SJ/SC WPCP is discharged through Artesian Slough and into South San Francisco Bay. During the summer months, about 10 percent (10-12 mgd) is recycled through South Bay Water Recycling pipelines for landscaping, agricultural irrigation, and industrial needs around the South Bay.

4.1.2 Feed Water Quality

The source of feed water for the pilot testing is tertiary treated wastewater. The tertiary treated wastewater is characterized by relatively high levels of TDS, hardness and alkalinity, with moderate levels of organic material and relatively low turbidity. Table 4-1 present the typical feed water quality at the pilot site.

Table 4-1. SJ/SC WPCP average pilot influent water quality

Parameter	Concentration	Unit
Cl^-	188	mg/L
NO_3-N	7.1	mg/L
SO_4	96	mg/L
Br^-	<1.0	mg/L
NO_2-N	<0.05	mg/L
Al	0.06	mg/L
Ba	0.020	mg/L
B	0.510	mg/L
Ca	59.1	mg/L
Cr (Total)	<0.002	mg/L
Fe	0.07	mg/L
Mg	31.7	mg/L
SiO_2	24.0	mg/L
Na	156	mg/L
Sr	0.387	mg/L
NH_3-N	<0.1	mg/L
Conductivity	1250	umhos/cm
pH	7.3	SU
TOC	9	mg/L
TKN	0.4	mg/L
TSS	<2	mg/L
Turbidity	0.7	NTU
Hardness, total (CaCO3)	250	mg/L
Alkalinity, total (CaCO3)	190	mg/L
TDS	750	mg/L
UV-254	0.109	Abs/cm
Fecal Coliforms	<1	MPN; cfu/ mL
Total Coliforms	1	MPN; cfu/ mL
Heterotrophic plate count	300	cfu /mL

4.2 General Pretreatment

To protect the membranes that are integral to the advanced treatment process investigated in this study, effluent from the San Jose/Santa Clara Water Pollution Control Plant was subjected to pretreatment prior to introduction to the reverse osmosis and electrodialyis units. Different types of chemicals were added to remove free chlorine that can dissolve the membranes, and various types of filters (including microfiltration) were used to remove suspended solids that can deposit on the membranes, increasing maintenance costs and reducing runtimes. These pretreatment processes are described briefly below.

4.2.1 Dechlorination

Tertiary treated water from the SJ/SC WPCP is disinfect with chlorine before being diverted to the SBWR transmission pumping station. Free chlorine concentrations observed in the effluent, which serves as the feed water to the pilot plant typically average 1-2 mg/L, with peaks of 4-8 mg/L. However, on occasion the concentration of free chlorine can reach 25 mg/L. Such spikes occur most often when recycled water demand drops rapidly after a period of high use, and are thought to result from delays in the automatic reduction of chlorination rates. As noted above, the non-thermal demineralization equipment used in this study (EDR and RO) includes sensitive membranes that can be damaged by prolonged exposure to free chlorine. The following are dechlorination strategies that were evaluated during pilot testing to reduce free chlorine and prevent irreversible membrane damage.

Granular Activated Carbon (GAC)

Two Ionics TurboFlo vessels were used for the removal of free chlorine in the feed water by reduction (Figure 4-2). Each 100 psi ASME code stamped carbon steel vessel had a capacity of 42 cubic feet. The GAC contactors were plumbed in series to ensure that, given the variability of free chlorine present in the feed water, complete dechlorination would be maintained to protect the membrane demineralization equipment. GAC contactors were used exclusively for the EDR pilot system.

Chloramination (Ammonia)

Harmful free chlorine concentrations were converted to chloramines. The ammonia feed pump was regulated and adjusted daily to ensure complete conversion of free chlorine to chloramines. Chloramination was used exclusively on the MF pretreatment equipment while polypropylene (free-chlorine sensitive) hollow fiber membranes were in use.

Sodium Bisulfite

Dechlorination was also achieved using sodium bisulfite. A dedicated pump was used to maintain a 4 to 5 mg/L dose of sodium bisulfite to chlorinated water entering the demineralization equipment. Sodium bisulfite was used as a pretreatment for both the RO and EDR pilot systems.

Figure 4-2. Ionics TurboFlo GAC contactors

Figure 4-3. Sodium bisulfite dechlorination feed pump and storage tank

4.2.2 Particulate Removal

For many membrane treatment processes, pretreatment is required to reduce suspended solids and colloidal matter for the prevention of membrane fouling. Suspended or undissolved matter in the feed water may deposit on the surface of the membrane as the water passes along or through the membrane. A build up of these deposits may eventually reduce the flow of water through the membrane and cause the applied pressure to increase. For both reverse osmosis and EDR systems, membrane fouling may, in part, contribute to a decrease in the salt or mineral rejection of the system, causing deterioration in the product water.

Conventional pretreatment

The use of prefilters as a pretreatment is common among all membrane systems to help prevent membrane fouling and minimize mechanical damage that may be caused by particulate matter. Prefiltration is typically accomplished by using cartridge filters (AWWA 1999). Cartridge filters (5-15 µm) were used as pretreatment for both RO and EDR processes.

Additionally, multimedia sand filtration (MMF) was pilot tested as a pretreatment to the EDR system. The pilot multimedia prefilter contained 10 cubic feet of filter media consisting of various sized, distinctly layered sand. The MMF could accommodate up to 100 gpm of flow with a maximum of 26 psi pressure loss.

Membrane Pretreatment

A major feature of current RO plants is the use of conventional pretreatment (Wilf 2001). Conventional pretreatment has several disadvantages including:

- lack of an absolute barrier to suspended particles and colloidal matter that can severely limit RO performance,
- fluctuation in feed water quality to RO,
- need for frequent backwashing of the filters used,
- biological growth in filters leading to RO membrane biofouling,
- chlorine used for biofouling control in filters can reach RO membrane and cause damage.

These operational issues lead to higher costs by forcing operation of RO systems at conservative operational parameters. Membrane pretreatment using MF or UF provides several advantages over conventional pretreatment. MF or UF can provide an absolute barrier to microorganisms, suspended particles, and colloids, leading to stable and high quality RO feed water. Consequently, the RO system can be operated at more aggressive operating conditions resulting in savings in both operational and overall costs.

<u>USFilter Memcor Pilot Equipment</u>

The USFilter Memcor CMF-S 16S10T pilot system tested included the following components (Figure 4-4):

- Feed Pump
- Up to sixteen membrane modules
- Air Compressor
- Data Logger

Figure 4-4. Memcor CMF-S 16S10T pilot unit

The pilot is equipped with a centrifugal pump and is run in direct filtration mode; that is, all feed passes through the membrane while filtering. An inlet feed valve, responding to level switches in the tank, controls the water level above the modules, so that they remained completely submerged. During filtration, water is drawn through the fiber walls (outside to inside) under suction. The microfiltered water is then directed to the filtrate tank or to service. All solids and particulate matter are removed from the feed water, and built on the outside of the fiber walls. A timer initiates regular backwashes after 30 minutes. The backwash uses air to scour the fibers, while a small amount of filtrate is pushed backwards through the fibers (inside to outside) to remove the fouling layer. The backwash duration is approximately 2.5 minutes. Dirty backwash water is drained away, and the tank is refilled prior to recommencing filtration.

The hollow fiber MF membrane modules can be arranged in four groups of four modules each (Figure 4-5). Polypropylene (PP) and polyvinylidene fluoride (PVdF) submerged modules, each with a nominal pore size of 0.1 micron, were evaluated during pilot testing. The membrane element specifications are presented in Table 4-2. The unit was equipped with an Allen-Bradley Co. PLC, pressure transmitters, flow meter, chlorine meter, conductivity meter and temperature measurement. The pressure transmitters monitored the transmembrane pressure (TMP, the driving force for filtration). Online instruments were connected to the PLC as well as a MEMLOG™ data logger. Control functions and data display were accessed via an Alan Bradley Panel view operator interface mounted on the front of the control panel.

Figure 4-5. CMF-S hollow fiber membrane modules

Table 4-2. CMF-S membrane module specifications

	Units	Value	Value
Manufacturer		US Filter	US Filter
Membrane Model and ID Number		119066 (for CMF-S)	119018 (for CMF-S)
Membrane Commercial Designation		S10V	S10T
Approximate Size of Element (length x diameter)	ft (m)	1.186 x 0.131 (3.892 x 0.433)	1.186 x 0.131 (3.892 x 0.433)
Active Membrane Area	ft^2 (m^2)	272 (25.3)	335 (31.09)
Number of Fibers per Module		9,600	14,500
Number of Modules (Operational)		9 in 16S10T pilot unit	11 in 16S10T pilot unit
Inside Diameter of Fiber	mm	0.5	0.39
Outside Diameter of Fiber	mm	0.8	0.65
Approximate Length of Fiber	m	1.1 m exposed length	1.1 m exposed length
Flow Direction		outside-in	outside-in
Nominal Membrane Pore Size	micron	0.1 um	0.1 um
Absolute Membrane Pore Size	micron	0.2 um	0.2 um
Membrane Material/Construction		Polyvinylidene Fluoride	Polypropylene
Membrane Surface Characteristics		hydrophobic	hydrophobic
Membrane Charge		neutral	neutral
Maximum Transmembrane Pressure	kPa (psig)	120 (17.4)	85 (12.3)
Acceptable Range of Operating pH Values		2 - 10	2 - 14
Acceptable Range of Operating Temperatures	degF (degC)	32 - 104 (0 - 40)	34 - 104 (0 - 40)
Chlorine/Oxidant Tolerance	ppm	200	<0.05

Chemical Consumption
Ammonia was added to the feed water at a dose of approximately 4 to 5 mg/L to eliminate free chlorine from harming polypropylene membranes. When PVdF membranes were used, no ammonia was necessary to reduce free chlorine. No chemicals were used during routine backwash, beyond any free chlorine or chloramines that might be present in the feed water to the PP or PVdF systems. CIP Chemical cleaning was periodically performed to remove foulants. CIPs were performed during this pilot testing with citric acid (10 lbs). The frequency of cleaning was determined as needed during the trial.

Waste Production
Backwash waste consists of water where the concentrations of naturally occurring particulates and organics are significantly higher (20 to 30 times) than raw water. Approximately 220 to 270 gallons of backwash waste were generated with each backwash. Cleaning chemical wastes consisted of approximately 140 to 170 gallons pH 2 to2.5 citric acid solution per cleaning.

4.3 Reverse Osmosis Equipment

The trial equipment consisted of a modified USFilter "H" series RO, model number ROSLH 3180 (Figure 4-6). The pilot system could utilize up to 12 vessels, but for the purposes of this study, 3 vessels configured in a 2:1 array with a minimum Feed Water Recovery (FWR) of 50 percent. It was originally proposed to test two different types of RO membranes. However, during commissioning of the pilot unit, the equipment vendor recommended evaluating two different pretreatment membranes due to the presence of variable free chlorine concentrations in the feed water (See Section 4.2). As a result the RO membranes elements evaluated were Dow Filmtec brackish water membranes, part number BW30-4040. Each vessel housed four RO membrane elements. The RO membrane element specifications are presented in Table 4-3.

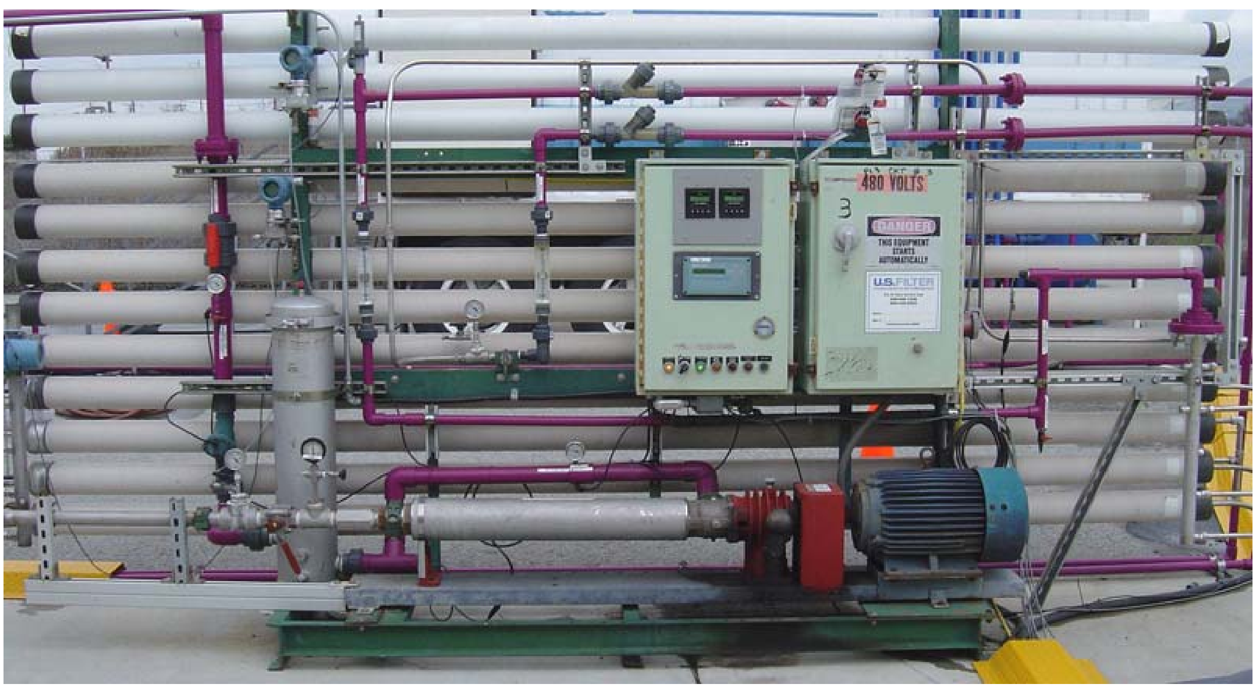

Figure 4-6. Reverse osmosis pilot system

Pretreatment
Several RO pretreatment options were available and used for pilot testing including:

- Microfiltration using Chlorine-Sensitive Polypropylene Membranes
- Microfiltration using Chlorine-Tolerant Polyvinylidene Fluoride Membranes
- Sodium Bisulfite
- Antiscalant (Argo 150; 1 to 2 mg/L)
- 5-15 µm Cartridge Filtration

Waste Production
The RO was operated at FWR between 50 and 65 percent. The maximum feed water flow was 20 gpm resulting in 5 to 10 gpm of concentrate generated by the RO process.

Table 4-3. RO membrane element specifications

	Units	Value
Manufacturer		Dow Filmtec
Membrane Model and ID Number		80783
Membrane Commercial Designation		BW30-4040
Approximate Size of Element (length x diameter)	length x diameter - in (mm)	40 x 3.9 (1016 x 99)
Active Membrane Area	ft^2 (m^2)	82 (7.6)
Number of Modules (Operational)		12 in ROSLH 3180 pilot unit
Applied Pressure	psig (bar)	225 (15.5)
Permeate Flow Rate	gpd (m3/d)	2,400 (9.1)
Stabilized Salt Rejection	%	99.5
Membrane Type		Polyamide Thin-Film Composite
Maximum Operating Temperature	ºF (ºC)	113 (45)
Maximum Operating Pressure	psi (bar)	600 (41)
Maximum Feed Flow Rate	gpm (m3/h)	16 (3.6)
Maximum Pressure Drop	psig (bar)	15 (1.0)
pH Range, Continuous Operation		2 to 11
pH Range, Short-Term Cleaning		1 to 12
Maximum Feed Silt Density Index		SDI 5
Free Chlorine Tolerance	ppm	< 0.1

4.4 Electrodialysis Reversal Equipment

The trial equipment consisted of an Aquamite V with a bipolar membrane stack. The capacity of the Aquamite V was 15,000 to 35,000 gpd. The maximum feed flow for this unit was 60,000 gpd. The Aquamite V supported an electric power supply of 480/460/380/220 Volts, 50/60 Hz, 3 phase and was supplied by direct current (DC) at 3 phases, full wave with silicon diode rectifiers.

As shown in Figure 4-7, the EDR pilot system was installed in Ionics' mobile pilot plant trailer. The trailer housed the EDR unit, control panel, multimedia filter, and cartridge filter. Located outside of the trailer were the two GAC filters and the sodium bisulfite feed pump and tank.

Figure 4-7. Electrodialysis reversal pilot plant

The EDR operated at 22 to 27 gpm to continually produce demineralized water without constant chemical addition during normal operation. Current was supplied at 2 to 4 amps depending on the specific water quality goals to be achieved. Membrane fouling and scaling was controlled by using electrical polarity reversal every fifteen minutes.

Typically, EDR is configured using multiple stages to provide the maximum membrane surface area and retention time to remove a specified fraction of salt from the demineralized stream. Two types of staging are used: hydraulic and electrical. For this study, the Aquamite V pilot unit operated as a single stack with two electrical stages that could be independently controlled to achieve a desired water quality. Electrical staging was accomplished by inserting additional electrode pairs into the membrane stack to provide maximum salt removal rates while avoiding polarization and hydraulic pressure limitations.

<u>Pretreatment</u>
Several pretreatment options were available for pilot testing prior to EDR treatment including:

- Granular Activated Carbon (GAC)
- Sodium Bisulfite
- Multimedia Sand Filtration
- Microfiltration (MF)
- 5-10 µm Cartridge Filtration

Although GAC and sodium bisulfite were used for dechlorination (to prevent damage of the membranes to due occasional high spikes of chlorine), it was determined that membrane biofouling due to algal growth could be controlled by maintaining a small amount of residual free chlorine to the EDR stack. This was achieved by bypassing a portion of the feed water flow to the dechlorination equipment and allowing it to enter the dechlorinated feed stream to the EDR stack. During this study, the average free and total chlorine concentration in the pilot feed water was 1 to 2 mg/L and 4 to 5 mg/L, respectively. Bypassing approximately 25% of the flow to the EDR allowed 0-0.5 mg/L free chlorine to be maintained in the EDR stack.

4.5 Water Quality

4.5.1 Analytical Methods

All off-site water quality analyses were performed at the City of San Jose's Environmental Services Department laboratory. Table 4-4 summarizes the methods used for all laboratory analyses performed.

Table 4-4. Summary of analytical procedures

Parameter	Method
Cl^-	EPA 300
NO_3-N	EPA 300
SO_4	EPA 300
Br^-	EPA 300
NO_2-N	EPA 354.1
Al	EPA 200.7
Ba	EPA 200.7
B	EPA 200.7
Ca	EPA 200.7
Cr (Total)	EPA 200.7
Fe	EPA 200.7
Mg	EPA 200.7
SiO_2	EPA 200.7
Na	EPA 200.7
Sr	EPA 200.7
NH_3-N	SM 4500-(NH3)H
Conductivity	SM 2510B
pH	SM 4500H+
TOC	SM 5310B
TKN	SM 4500 N(org)-C
TSS	SM 2540D
Turbidity	SM 2130B
Hardness, total (CaCO3)	SM 2340C
Alkalinity, total (CaCO3)	SM 2320B
TDS	SM 2540C
UV-254	SM 5910B
Fecal Coliforms	SM 9221A/ 9222A
Total Coliforms	SM 9221D/ 9222D
Heterotrophic Plate Count	SM 9215

4.5.2 Sampling Protocol/Frequency

All water quality samples were collected as grab samples using sample containers provided from the corresponding laboratory. All samples were transported to the lab in a cooler and were processed within the allowable holding period. During sampling, sample ports were allowed to flush before samples were collected.

4.5.3 Quality Assurance / Quality Control

The following QA/QC procedures were followed during pilot testing.

Pilot Plants Auxiliary Units

The pilot plant auxiliary equipment such as electronic pressure sensors, flow meters, volt and amperage meters, and safety switches were not calibrated on-site during the pilot testing start up period as outlined in the Advance Water Treatment Pilot Study Work Plan. Calibrations of selected equipment occurred during the testing period.

On-line Monitoring Devices

The readings from on-line pH meters, conductivity meters and thermocouples were verified by comparison to grab samples collected, submitted and analyzed by the City of San Jose Environmental Services Department laboratory.

Data Analyses

Data collected on-site was regularly merged with data obtained from off-site laboratory analyses to form a comprehensive database for data analyses, retrieval, reporting, and graphics. A modular database program was developed for this project to include all produced data. All data was checked and verified by the project engineer before and after entry into the database program.

5. Results and Discussion

5.1 Reverse Osmosis with Membrane Pretreatment

Membrane (microfiltration) pretreated water was continuously fed to the RO pilot system at a flow rate of approximately 20 gpm. The pilot system utilized DOW FilmTec BW30-4040 membrane elements. Four elements were placed in series in each of the three pressure vessels configured as a two-stage, 2:1 array. Sodium bisulfite and antiscalant were added to the MF pretreated water to control RO membrane fouling and protect the membrane elements from chemical damage due to free chlorine or chloramines.

Two membrane pretreatment scenarios were evaluated. First, during Phase I, polypropylene (PP) hollow fiber membranes were used in the CMF-S pilot system. However, because PP membranes are easily damaged by free chlorine, ammonia was added to the pilot plant feed water to create chloramines. After MF pretreatment, sodium bisulfite was added to dechlorinate the RO feed water. During Phase II, the PP MF membranes were replaced with chlorine-tolerant polyvinylidene fluoride (PVdF) membranes. The ammonia feed was removed and sodium bisulfite was used to dechlorinate the feed to the RO pilot.

5.1.1 Phase I – MF Pretreatment using PP Membranes

A process flow schematic of the Phase I pilot operation is provided in Figure 5-1. As shown, tertiary treated wastewater from the SJ/SC WPCP was dosed with 4-5 mg/L ammonia to form chloramines and fed to the USFilter CMF-S. The CMF-S operated at a total flow rate of 45 gpm using 11 PP hollow fiber membrane modules (approximately 4 gpm per module). At this flow rate, the operating flux was 22 gfd. A portion of the MF permeate was stored in a separate backwash tank and used for regular backwash of the membranes every 30 minutes. The remaining MF permeate was stored in a large break tank downstream, which served as the feed to the RO system.

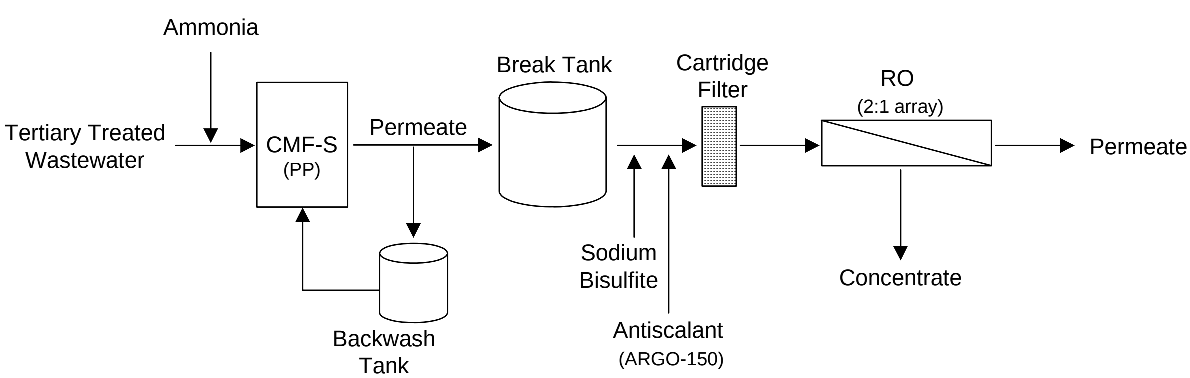

Figure 5-1. Reverse osmosis pilot testing schematic (PP membrane pretreatment)

The main objective of membrane pretreatment is to remove harmful suspended solids and colloidal matter that could damage the RO membranes. The silt density index (SDI) was measured both on the feed and permeate of the CMF-S to characterize the fouling potential of the RO feed water. As shown in Figure 5-2, the average SDI in the tertiary treated wastewater was 7.0. It is important to note that, certain feed samples caused the 0.45 µm filter used for SDI measurement to become plugged within 15 minutes. In those instances, a modified SDI was calculated by determining the time to 100 percent pluggage. After membrane pretreatment, the SDI was maintained at an average of 1.0. The maximum SDI allowed to the RO membranes, as recommneded by the manufacturer, is SDI 5.

The CMF-S was operated continuously for 1100 hours at a 20°C temperature-corrected flux of approximately 22 gfd, as shown in Figure 5-3. This flux was conservative and recommended by the equipment vendor. The transmembrane pressure was maintained between 1-1.5 psi during the first 450 hours of operation (Figure 5-4) and then increased to 3 psi during the remainder of operation. The increase in TMP was due to the fouling of the PP membranes over time. The resulting drop in the membrane permeability or specific flux can be seen Figure 5-3 for the last 650 hours of operation. Despite the minimal fouling trend observed during this test period, the CMF-S was able to continually remove suspended solids in the raw feed water, making it suitable to be subsequently treated using RO membranes. Additionally, it is generally expected that performing a CIP chemical cleaning at the end of this operational period could recover the membrane permeability. Unfortunately, membrane performance after a CIP cleaning was not evaluated due to failure of the ammonia feed system, causing the PP membrane fibers to be damaged by free chlorine exposure.

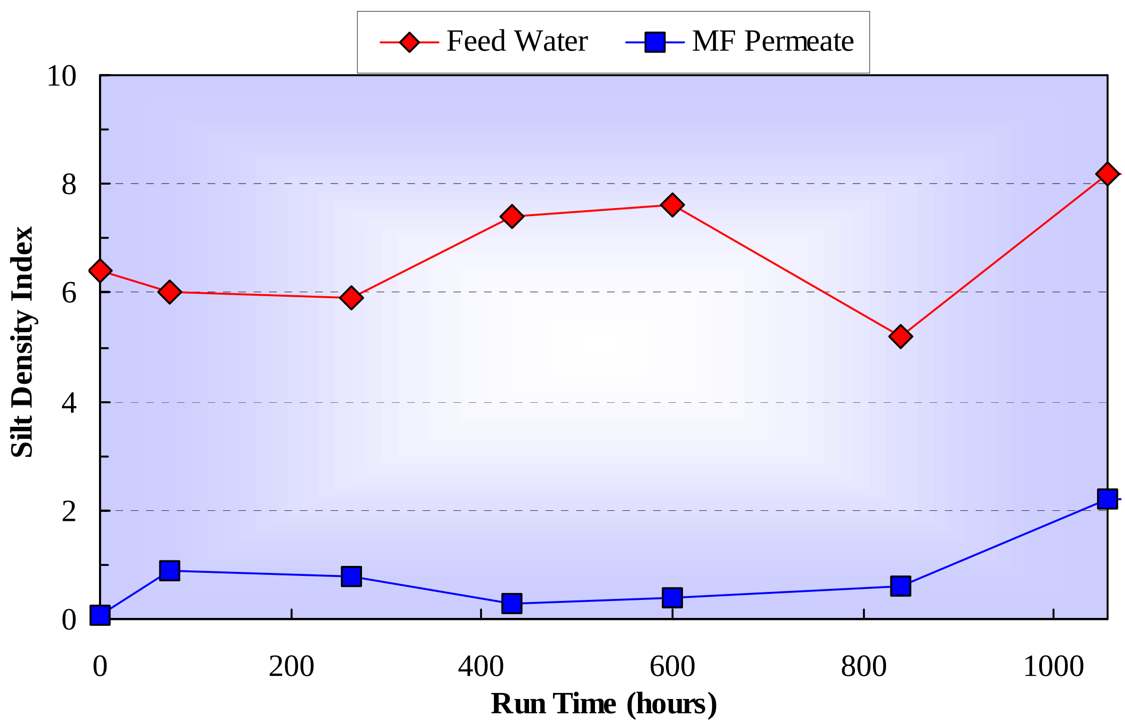

Figure 5-2. Silt density index (SDI) measurements – Phase I

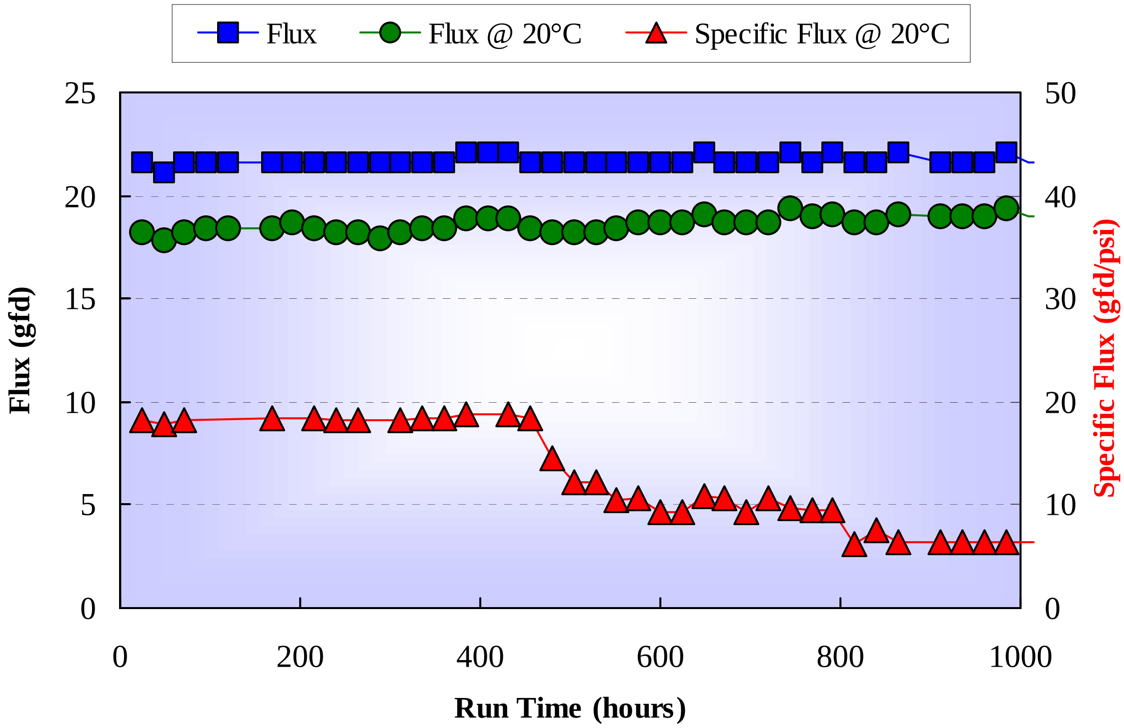

Figure 5-3. CMF-S temperature-corrected operating flux and specific flux – Phase I

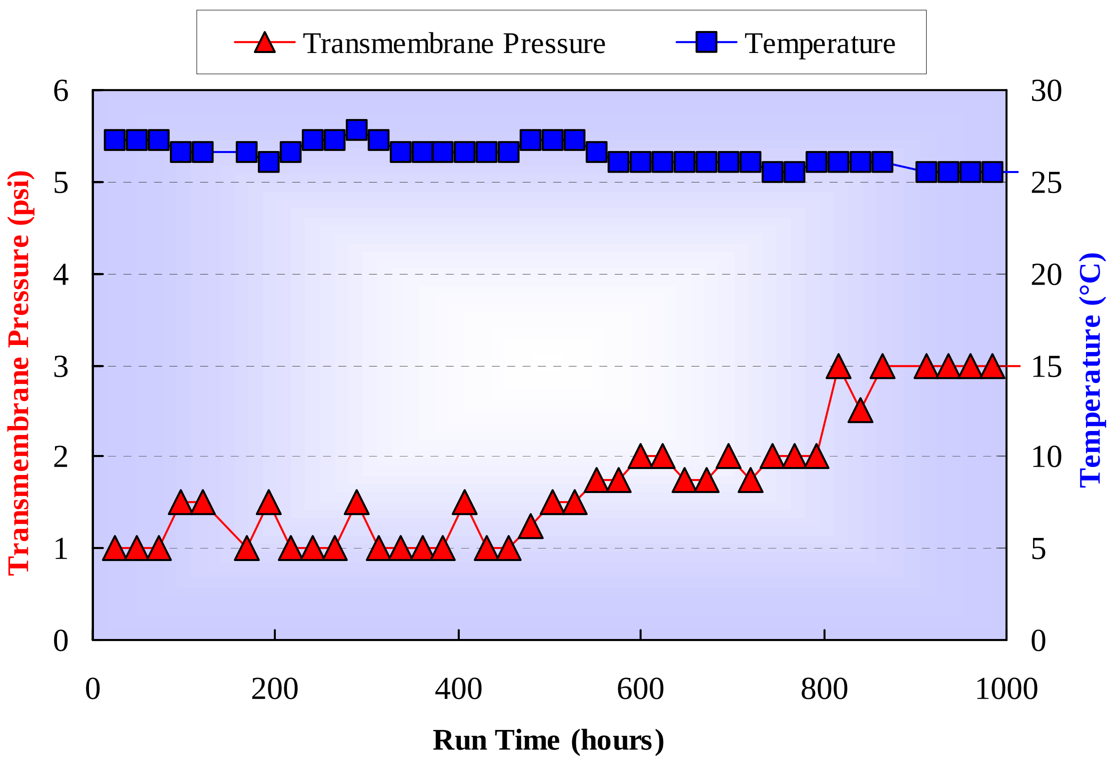

Figure 5-4. CMF-S transmembrane pressure and temperature – Phase I

Using MF pretreated water, the RO system operated continuously for 1100 hours. A summary of the RO operation and performance data is presented in Table 5-1. RO feed water (MF permeate) was dosed with 4-5 mg/L sodium bisulfite and 1-2 mg/L antiscalant (ARGO-150). Before entering the first stage of the RO, a 5-µm cartridge prefilter was used to remove fine particulate matter that might enter the system as a result of a failure in the pretreatment system.

Table 5-1. Summary of RO operation and performance – Phase I

Parameter	Range	Average
Feed Water Flow Rate (gpm)	19.3 – 23	20.7
Product Flow Rate (gpm)	10 – 11.5	10.6
Feed Water Recovery (%)	48 – 55	51
Flux @ 25°C (gfd)	14.5 – 17	15
Specific Flux @ 25°C (gfd/psi)	0.07 – 0.09	0.08
Feed Water Temperature (ºC)	25 – 27	26.5
Feed Water Pressure (psi)	215 – 245	230
Feed TDS (mg/L)	660 – 795	720
Product TDS (mg/L)	6 – 10	7
Salt Rejection (%)	98.5 – 99	99
Feed Water SDI	0.1 – 2	0.9

Variations in performance due to temperature fluctuations were negligible for the testing period as the RO influent water temperature averaged 26°C. As shown in Figure 5-5, the RO operated at a temperature corrected flux of 15 gfd. Fed at a flow rate of 20 gpm, the RO operated at a FWR of 50 percent (Figure 5-6) producing 10 gpm of permeate while the remaining 10 gpm was disposed of as concentrate. The calculated specific flux based on the net driving pressure (NDP) was 0.07 to 0.08 gfd/psi. To achieve the high rejection of salts, the RO was operated with a net driving pressure of approximately 200 psi (Figure 5-7). Despite some slight variation in the operating pressure the NDP was maintained throughout the testing period without significant increase or loss.

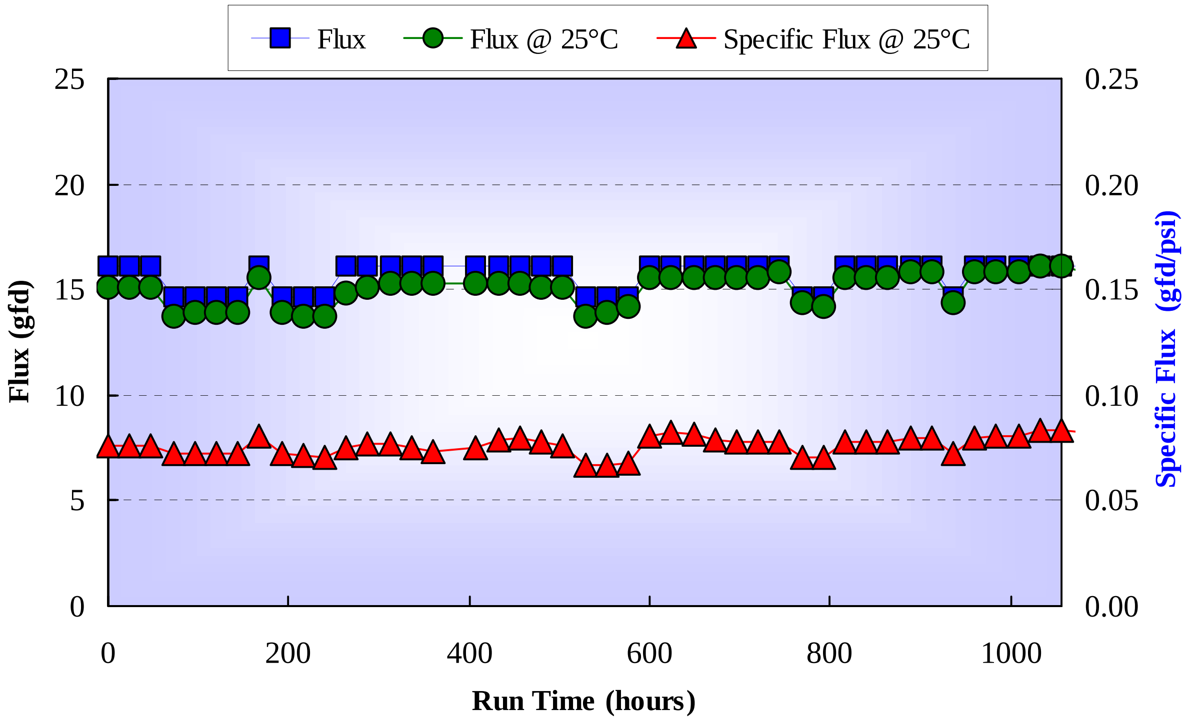

Figure 5-5. RO temperature-corrected operating flux and specific flux– Phase I

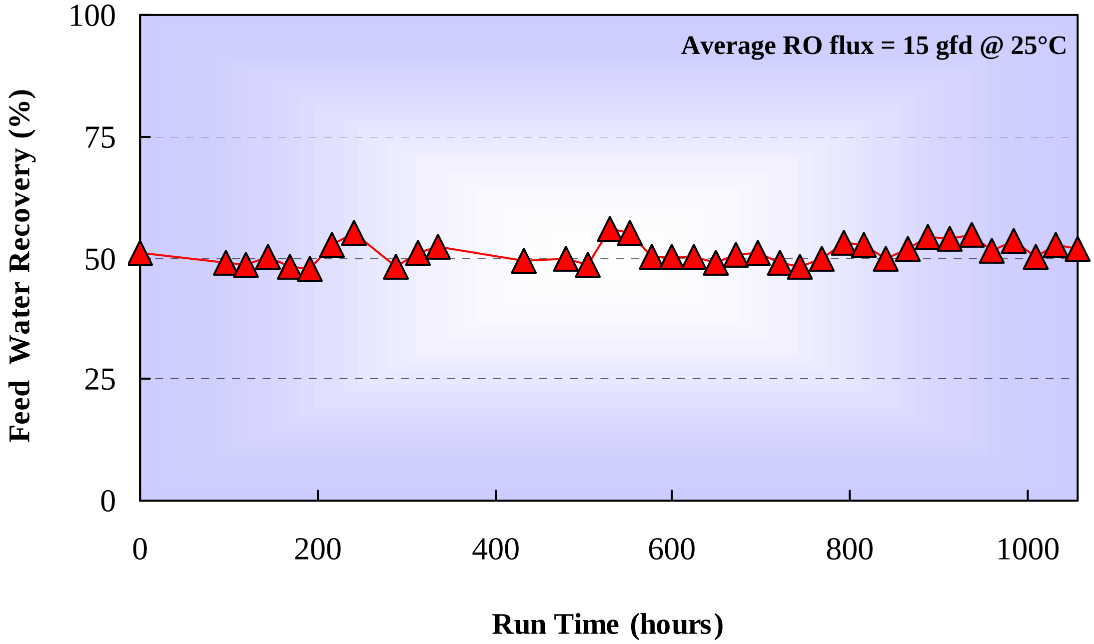

Figure 5-6. RO feed water recovery – Phase I

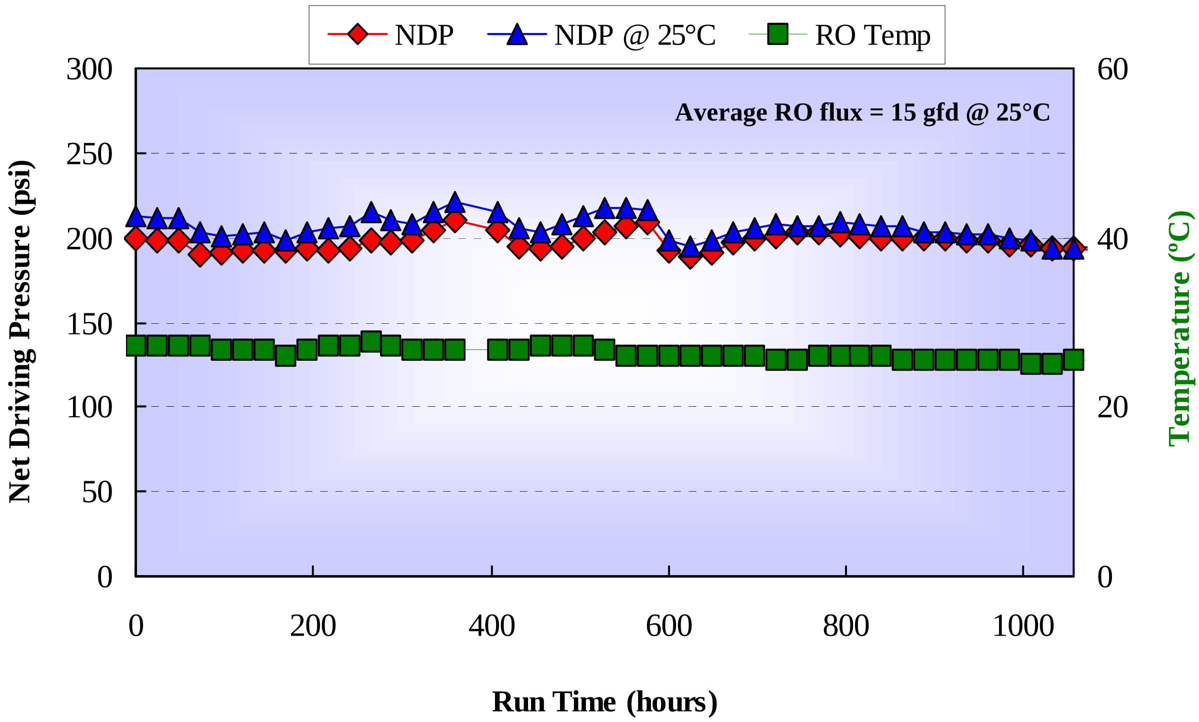

Figure 5-7. RO net driving pressure – Phase I

Up to 99 percent TDS rejection (1 percent salt passage) was achieved under these operating conditions, as shown in Figure 5-8. The feed water TDS averaged 720 mg/L and the permeate TDS averaged 7 mg/L (Figure 5-9). Although the overall RO permeate quality was not affected, it is interesting to note that the influent TDS increased and decreased on a weekly basis. This may have been due to regular operation and/or the demand experience by the SJ/SC WPCP. Table 5-2 summarizes the amount of salt rejected for specific ions by the RO.

Table 5-2. Average RO salt rejections

Ion	Feed (mg/L)	Product (mg/L)
Chloride	200	1.4
Nitrate as N	8	0.2
Sulfate	100	<4
Calcium	54	0.3
Magnesium	31	0.1
Silica	25	0.4
Sodium	150	3.5
Conductivity (µS)	1230	12
TDS (mg/L)	720	7

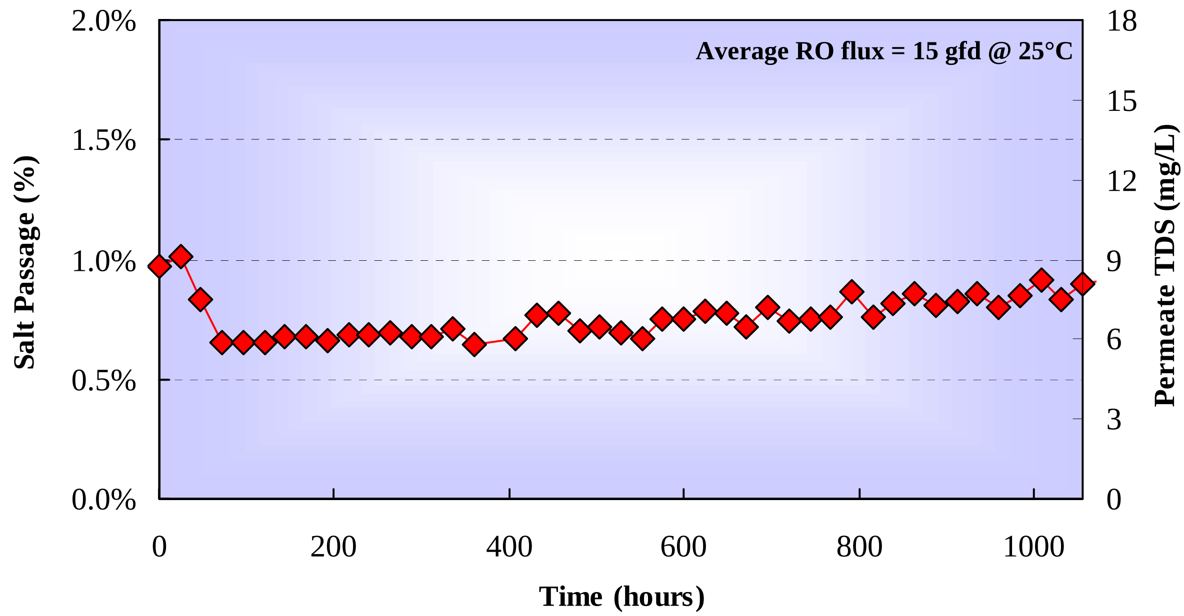

Figure 5-8. RO salt passage – Phase I

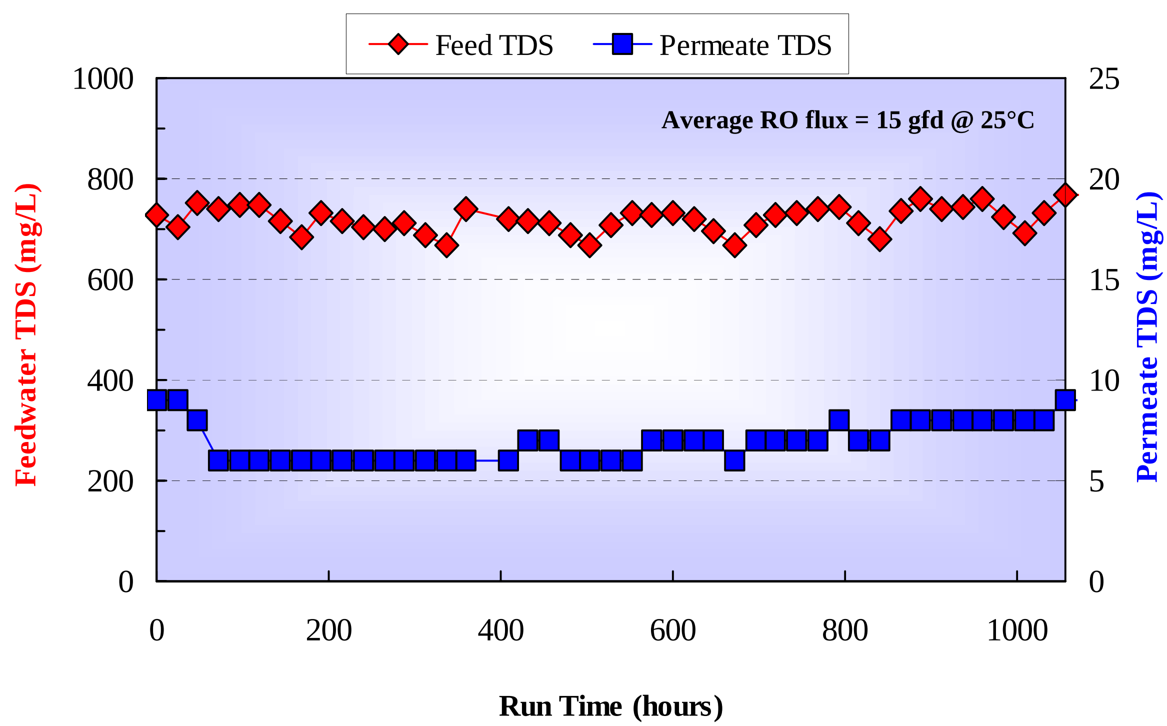

Figure 5-9. RO total dissolved solids (TDS) levels – Phase I

As previously discussed, the ammonia feed pump for the CMF-S feed water failed after 1100 hours of continuous operation. As a result, free chlorine in the pilot plant feed water came into contact with the chlorine-sensitive polypropylene membranes and caused the CMF-S hollow fibers to break. Concurrent, but unrelated to this event, RO performance also decreased and it was quickly discovered that the O-ring seals used in the RO vessels had become worn causing poor water quality and damaged RO membrane elements.

This event concluded Phase I testing and encouraged the Project Team to explore a more robust option to protect RO membranes from harmful damage. After discussion with the USFilter, it was recommended that chlorine-tolerant PVdF membranes be used in this reclaimed water application to ensure the performance of the RO process. Additionally, replacement RO membrane elements were provided and used for additional testing.

5.1.2 Phase II – MF Pretreatment using PVdF Membranes

The objective of Phase II testing was to operate of the RO membranes using MF pretreatment with chlorine-tolerant PVdF membranes. A process flow schematic of the Phase II pilot operation is provided in Figure 5-10. Tertiary treated wastewater from the SJ/SC WPCP, with an average free chlorine residual of 1 mg/L, was fed directly to the USFilter CMF-S system. Dechlorination, however, was not necessary since PVdF hollow fiber membrane modules were used.

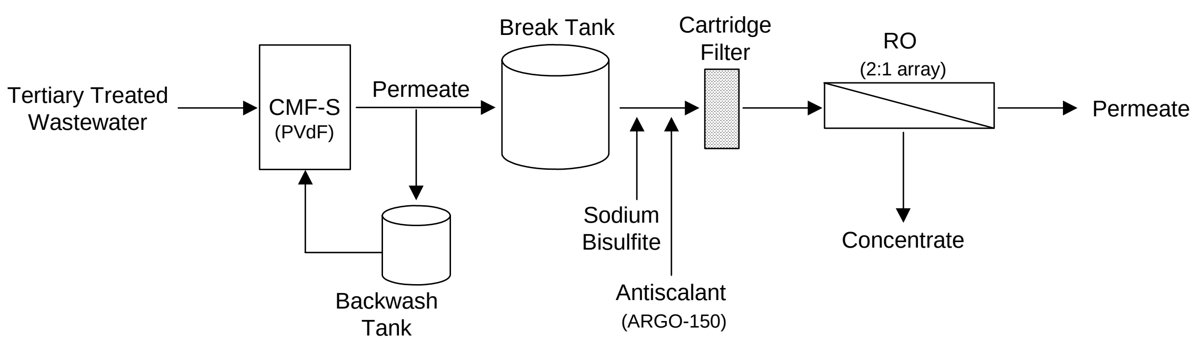

Figure 5-10. Reverse osmosis pilot testing schematic (PVdF membrane pretreatment)

As in Phase I, the silt density index (SDI) was monitored weekly for both the feed and permeate of the CMF-S to characterize the fouling potential of the RO feed water. As shown in Figure 5-11, the modified SDI in the tertiary treated wastewater averaged 13.0. After membrane pretreatment, the SDI averaged 0.3.

The CMF-S operated at a flow rate of 45 gpm for 500 hours with an applied flux of 30 gfd (specific flux = 8 to 10 gfd/psi) using 8 PVdF membrane modules (Figure 5-12). As shown in Figure 5-13, the transmembrane pressure was stable throughout all conditions and no significant membrane fouling was observed. It is important to note that no fatal flaws in the membrane pretreatment scheme were observed when operated with PVdF membranes on reclaimed wastewater from the SJ/SC WPCP. Typically, reclaimed wastewater applications have exclusively used polypropylene membranes. These test results represent one of the first pilot trials in which PVdF membranes were utilized for the pretreatment of RO feed water in a reclaimed wastewater application. Overall, PVdF performed as well as the PP membranes, and at a higher operating flux. An additional benefit was that the membranes were protected from free chlorine and any temporary spikes that may occur.

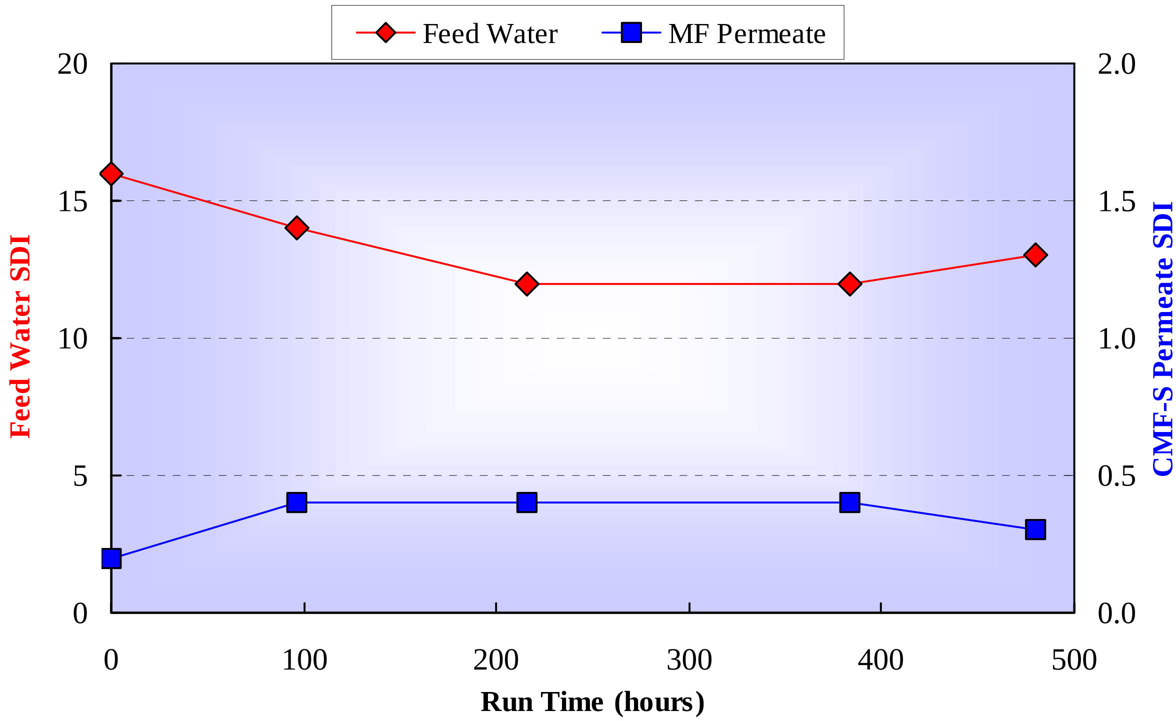

Figure 5-11. Silt density index (SDI) measurements – Phase II

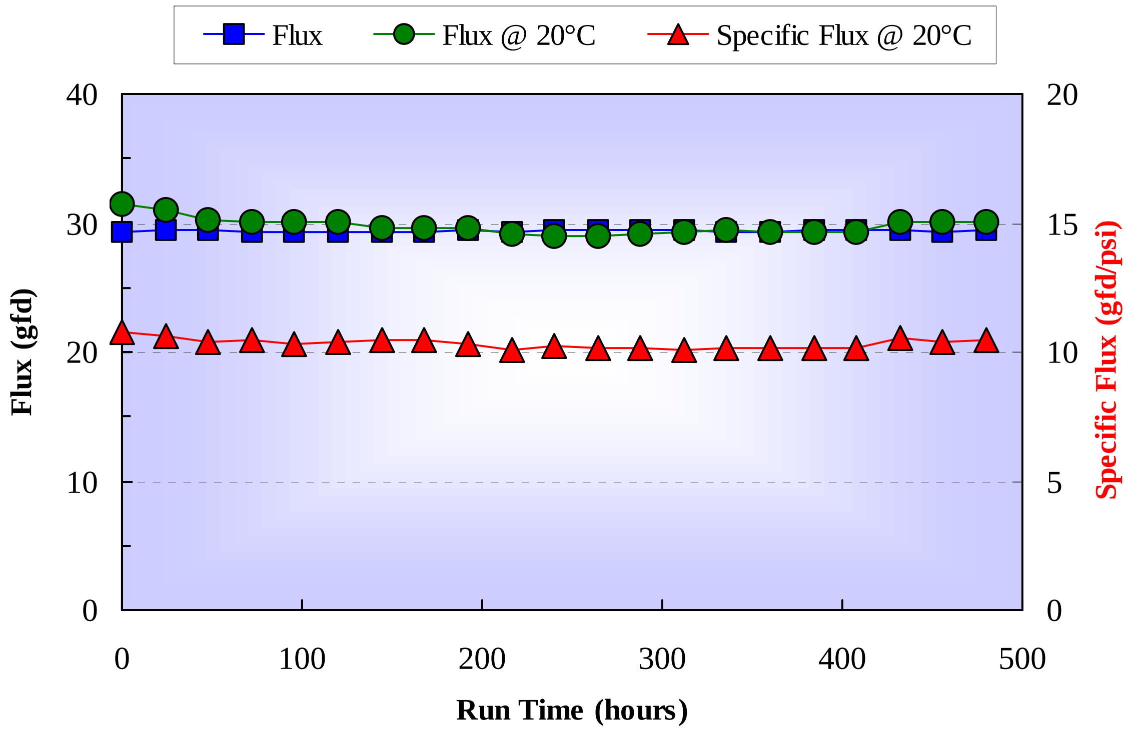

Figure 5-12. CMF-S temperature-corrected operating flux and specific flux – Phase II

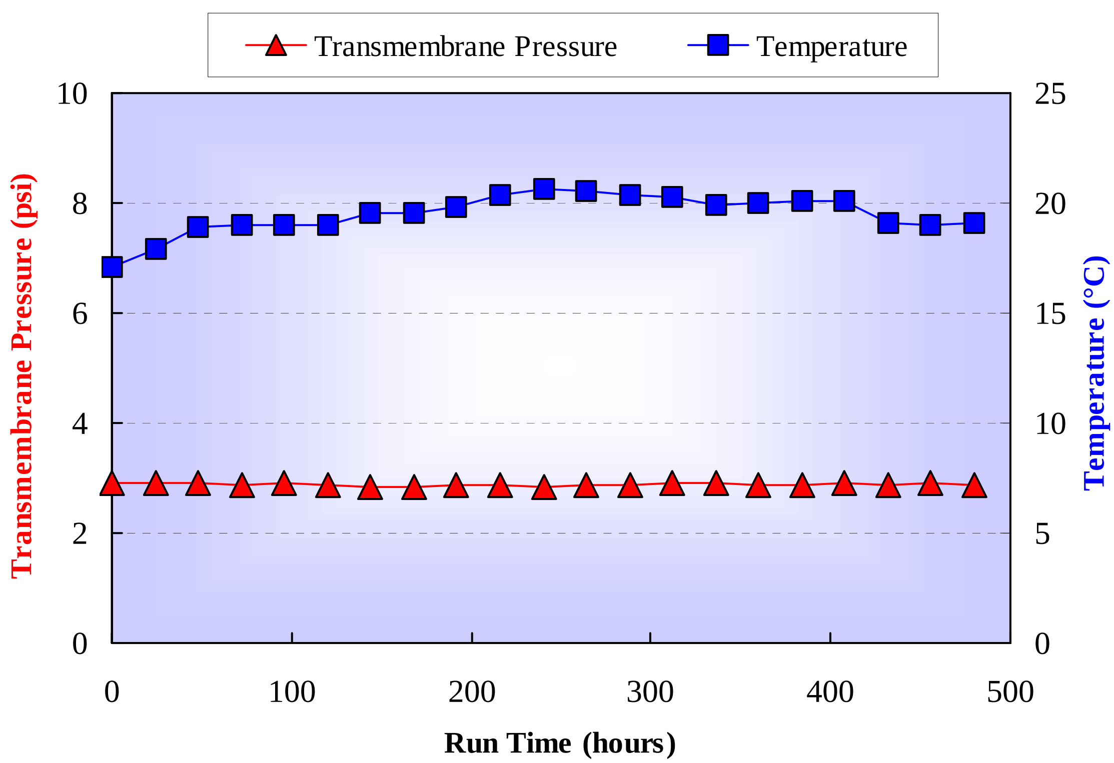

Figure 5-13. CMF-S transmembrane pressure and temperature – Phase II

A summary of the RO operation, performance data, and salt rejection of specific ions is presented in Table 5-3. The RO system was operated continuously for 500 hours at an applied flux of 15 gfd @ 25°C and an average FWR of 65 percent (Figure 5-14 and Figure 5-15). CMF-S pretreated reclaimed wastewater was dosed with 4-5 mg/L sodium bisulfite and 1-2 mg/L antiscalant (ARGO-150) and fed to the RO at 14 gpm. At a FWR of 65 percent, 9 gpm of RO permeate was produced. Before entering the first stage of the RO, a 5-µm cartridge prefilter was used to remove any additional fine particulate matter.

Table 5-3. Summary of RO operation and performance – Phase II

Parameter	Range	Average
Feed Water Flow Rate (gpm)	13 – 15	14
Product Flow Rate (gpm)	8 – 9.5	9
Feed Water Recovery (%)	60 – 69	65
Operation Flux (gfd)	14.5 – 15	15
Specific Flux (gfd/psi)	0.14 – 0.17	0.15
Feed Water Temperature (ºC)	21 – 22	22
Feed Water Pressure (psi)	100 – 120	102
Feed TDS (mg/L)	640 – 760	720
Product TDS (mg/L)	16 – 30	20
Salt Rejection (%)	96 – 98	98
Feed Water SDI	0.2 – 0.4	0.3

The NDP (corrected to 25ºC) of less than 90 psi (Figure 5-16) was measured during operation with new membranes, compared to 200 psi previously observed during Phase I. The lower NDP required during Phase II were most likely due to the fact that new RO membranes were used to replace the fouled RO membranes. Discussions with USFilter revealed that the original RO elements used during Phase I testing were refurbished membranes that had been used before. The unknown difference in conditions of the PP (Phase I) membranes to the PVdF (Phase II) membranes, may explain the large improvement in NDP during Phase II.

A slight increase in the NDP was observed throughout the test period from an initial 80 to 90 psi. The pressure increase may have been due to slow fouling over time. Additional long-term pilot testing would be required at this flux to determine the point at which the potential fouling of the membranes would result in diminished product water quality.

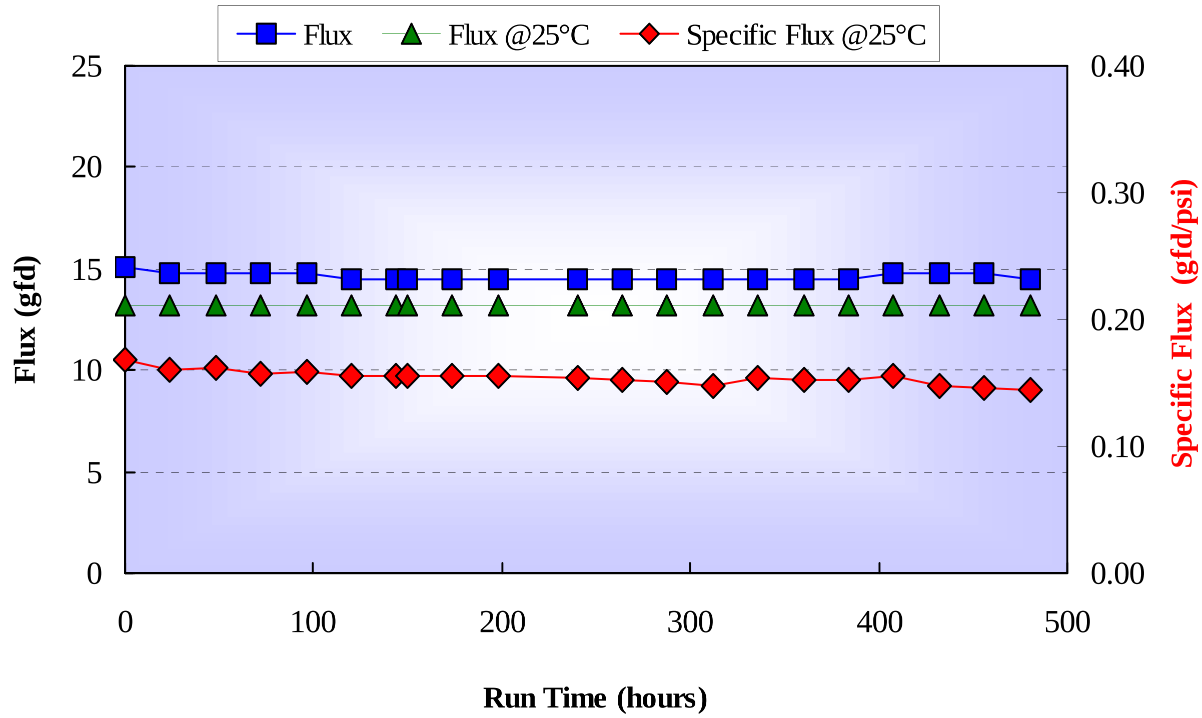

Figure 5-14. RO temperature corrected operating flux and specific flux– Phase II

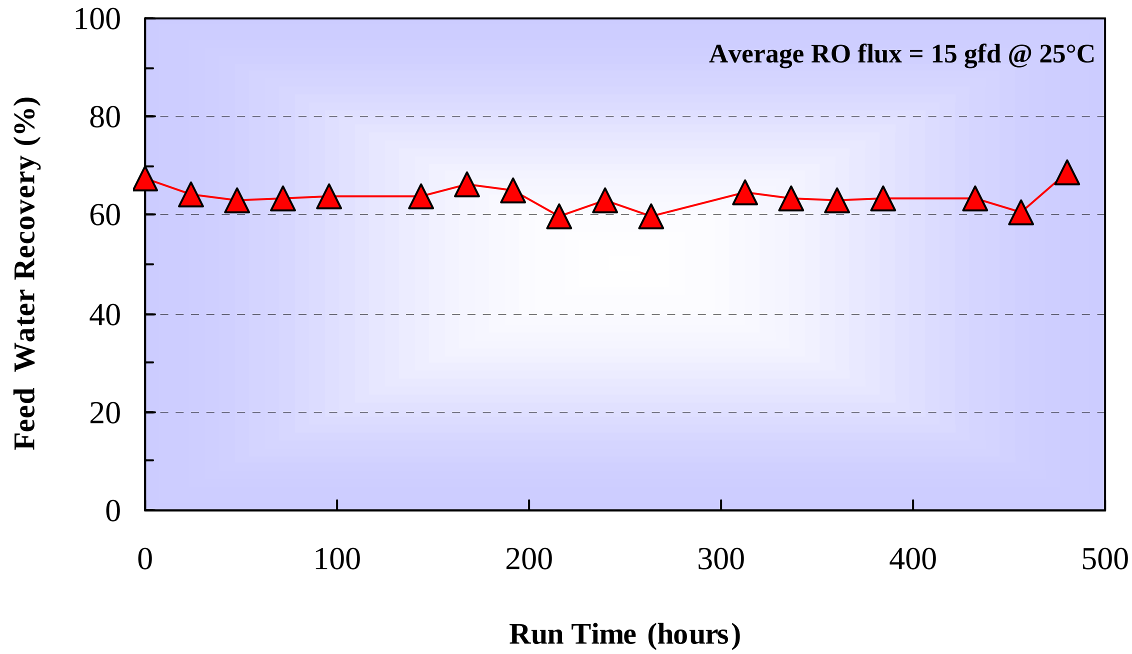

Figure 5-15. RO feed water recovery – Phase II

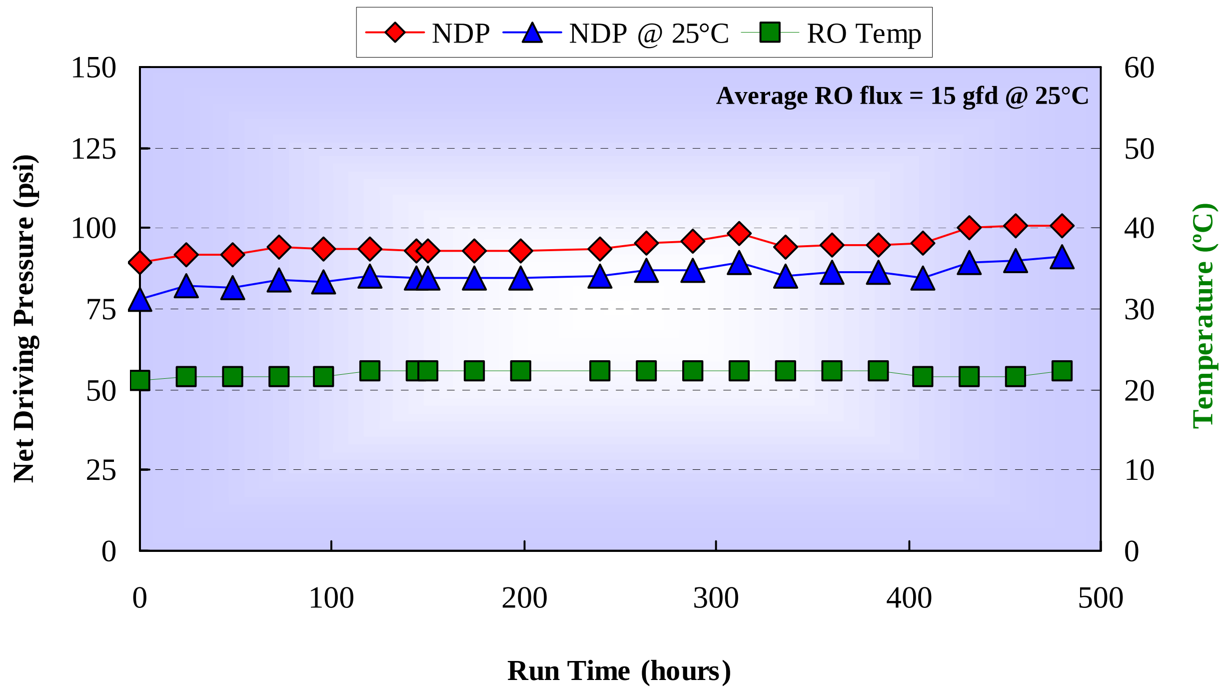

Figure 5-16. RO net driving pressure – Phase II

Up to 98 percent TDS rejection (2 percent salt passage) was achieved under these operating conditions, as shown in Figure 5-17. The feed water TDS averaged 720 mg/L and the permeate TDS averaged 20 mg/L (Figure 5-18). Table 5-4 summarizes the amount of salt rejected for specific ions by the RO.

Table 5-4. RO salt rejections – Phase II

Ion	Feed (mg/L)	Product (mg/L)
Chloride	191	13
Nitrate as N	9.2	1.2
Sulfate	119	<4
Calcium	52	0.6
Magnesium	31	0.4
Silica	25	3.5
Sodium	155	14.5
Conductivity (µS)	1250	33
TDS (mg/L)	720	20

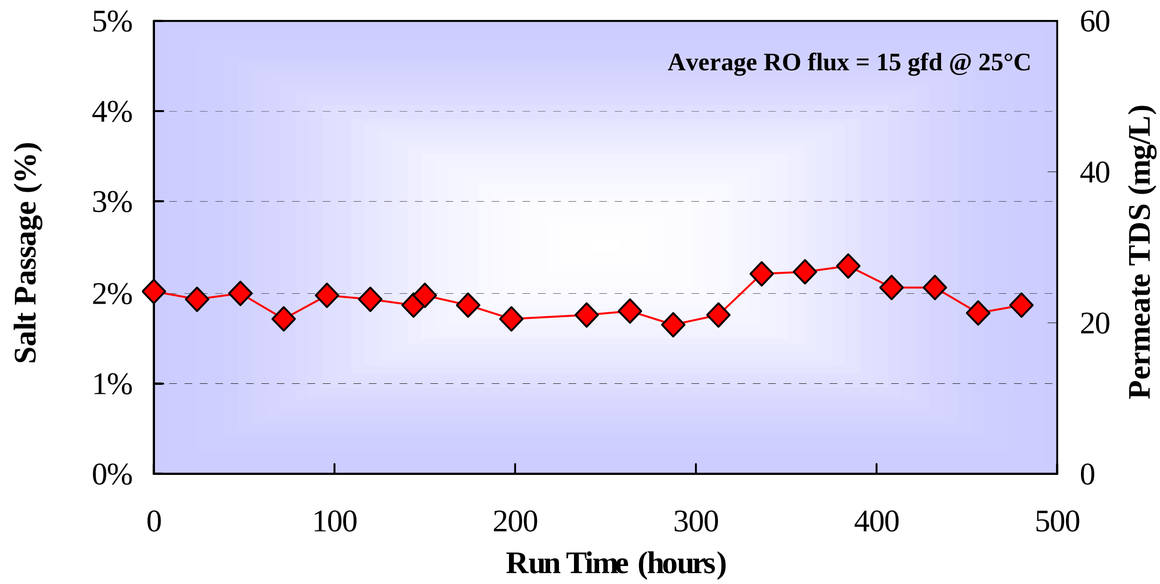

Figure 5-17. RO salt passage and permeate TDS – Phase II

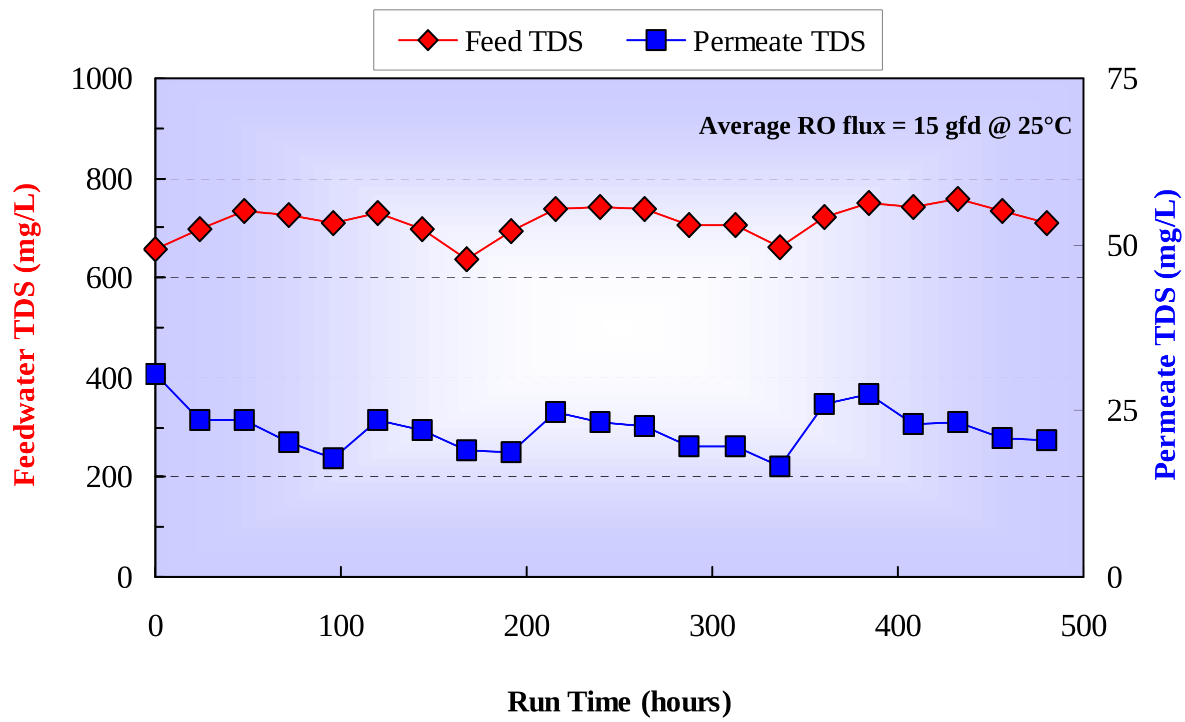

Figure 5-18. RO total dissolved solids (TDS) levels – Phase II

5.2 Electrodialysis Reversal

The two primary objectives of the EDR pilot testing were to:

- Evaluate system performance of the EDR to produce 350 and 500 mg/L TDS water, and
- Assess the impact of alternative pretreatment methods for removal of suspended solids [Granular Activated Carbon (GAC)/Multimedia Filtration (MMF), Microfiltration (MF) and, Cartridge Filtration only]

The first phase of testing included operating the EDR pilot to remove reduce the influent TDS to the 350 mg/L target treatment goal (baseline performance). Baseline performance was established using the manufacturer's recommended pretreatment scheme: this included GAC/MMF. Once baseline performance had been established, the EDR was adjusted to produce product water with 500 mg/L TDS to establish performance and identify any potential cost savings due to anticipated lower energy consumption. Additionally, during this second phase of operation, the three different types of pretreatment options (GAC/MMF, MF, Cartridge Filtration only) were evaluated.

5.2.1 Phase I – Baseline Operation (350 mg/L TDS Treatment Goal)

A general process flow schematic of the EDR pilot operation is provided in Figure 5-19. As shown, tertiary treated wastewater from the SJ/SC WPCP was dechlorinated by two GAC contactors in series. Suspended solids removal was achieved using a multimedia sand filter and a 5-10 µm cartridge filter. To prevent biological fouling on the EDR membranes, a small amount of the chlorinated feed water was bypassed to maintain a 0 to 0.5 mg/L free chlorine residual entering the EDR stack.

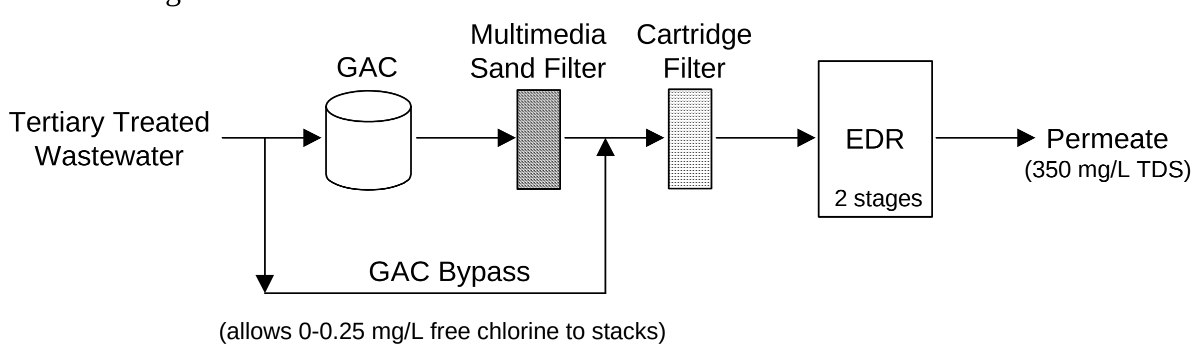

Figure 5-19. EDR pilot testing schematic (Phase I)

In Phase I, the EDR pilot operated continuously for 1050 hours at a feed flow rate of 28 gpm (Figure 5-20). Table 5-5 summarizes the operation and performance of the EDR in this test period. Approximately 28 gpm of feed water was provided to produce 24 gpm of demineralized water, resulting in an 85 percent feed water recovery. The pilot unit utilized a single stack with two electrical stages to reduce the influent TDS of 750 mg/L to 350 mg/L. Figure 5-21 demonstrates that a consistent effluent water quality was maintained throughout the test period and Table 5-5 summarizes the amount of salt rejected for specific ions by the EDR.

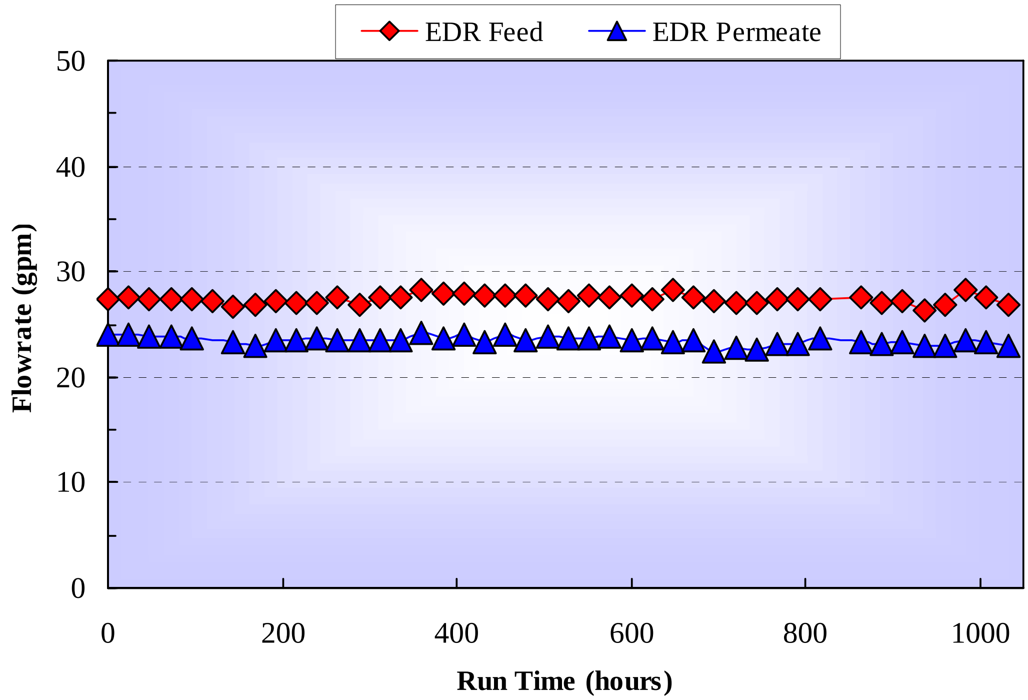

Figure 5-20. EDR pilot flowrates – Phase I

Table 5-5. Summary of baseline EDR operation and performance

Parameter	Range	Average
Feed Water Flow Rate (gpm)	26 - 28	27.5
Product Flow Rate (gpm)	22.5 – 24	23.5
Feed Water Recovery (%)	83 - 88	85.5
Feed Water Temperature (ºC)	25 - 27	26.5
Feed TDS (mg/L)	660 - 750	720
Product TDS (mg/L)	330 - 390	360
Stage 1 Voltage (V)	53 - 54	53.3
Stage 1 Current (amps)	3.3 – 4.3	3.8
Stage 2 Voltage (V)	49 – 51	50
Stage 2 Current (amps)	3 – 3.7	3.3
Stack Inlet Pressure (psi)	24 - 34	28
Stack Inlet DP (inches H_2O)	74 - 100	87.5
Stack Outlet Pressure (psi)	5 - 8	6.5
Stack Outlet DP (inches H_2O)	28 - 52	40
Salt Rejection (%)	53 - 57	53

Table 5-6. Summary of baseline EDR water quality

Ion	Feed (mg/L)	Product (mg/L)
Chloride	200	88
Nitrate as N	8	3.9
Sulfate	100	36
Calcium	54	12
Magnesium	31	8
Silica	25	23
Sodium	150	106
Conductivity (µS)	1230	650
TDS (mg/L)	750	350

The pressure drop through the membrane stack is dependent upon the spacer type, flow per stage, and number of pairs in each stage, and may also increase as a result of membrane fouling. As specified by the equipment vendor, the Aquamite V pilot should operate such that the stack inlet pressure does not exceed 50 psi. As shown in Figure 5-22, the stack inlet pressure was maintained below the 50 psi limit. However, during the testing period, the inlet pressure increased from 25 to 35 psi. This may be indicative of fouling of the membrane stack and could be related to the slight increase in both product TDS and electrical resistance that was observed after 400 hours of run time. The stack outlet pressure was measured to be approximately 6 psi throughout the testing period.

In order to achieve TDS removal to 350 mg/L, approximately 4 amps/53 volts and 2.8 amps/48 volts were applied to the first and second electrical stages, respectively. The voltage and amperage were monitored throughout the testing period and ensured that a constant electric potential was supplied to the EDR. Membrane fouling may cause a decrease in the applied current (at a constant voltage) and as a result, will cause the electrical resistance to increase. As seen in Figure 5-23, a slight increase in the electrical resistance was observed after 400 hours, which may have indicated that slight membrane fouling was occurring.

Despite the slight increase in both stack pressure and resistance, the EDR was able to consistently meet the water quality goal during the pilot test program. The observed fouling rate was minimal and is characteristic of normal operation.

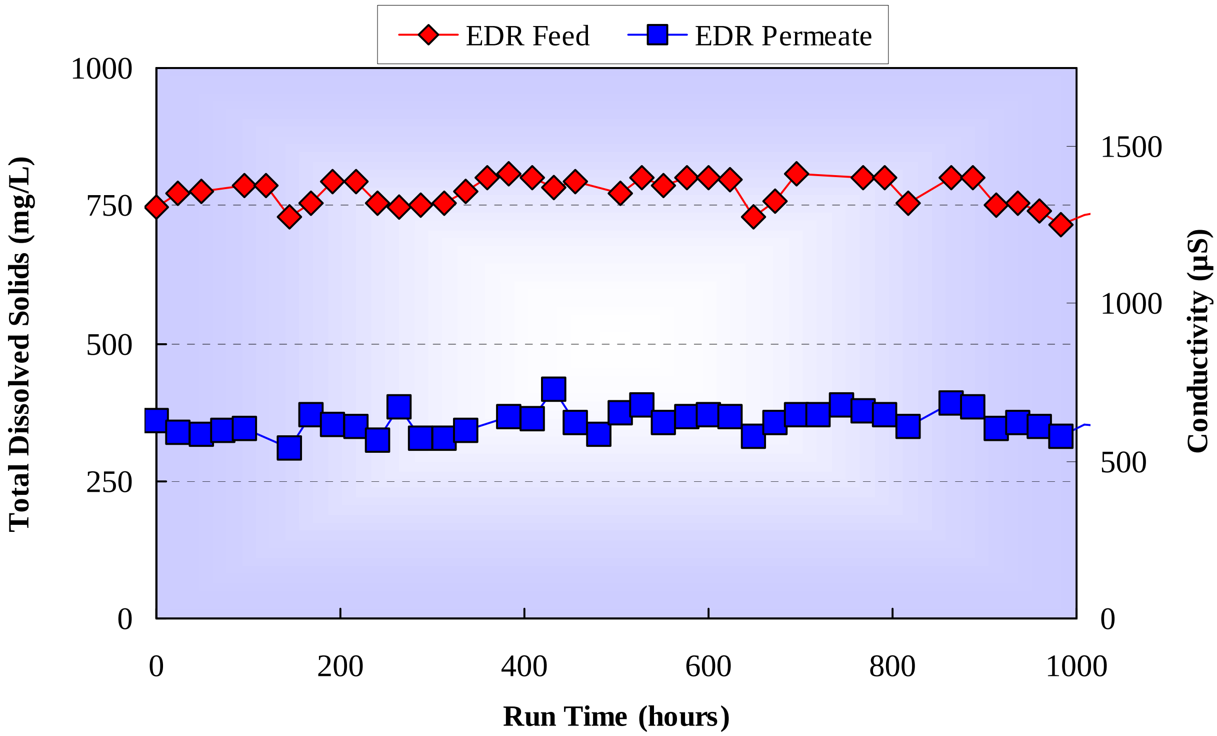

Figure 5-21. EDR pilot TDS levels – Phase I

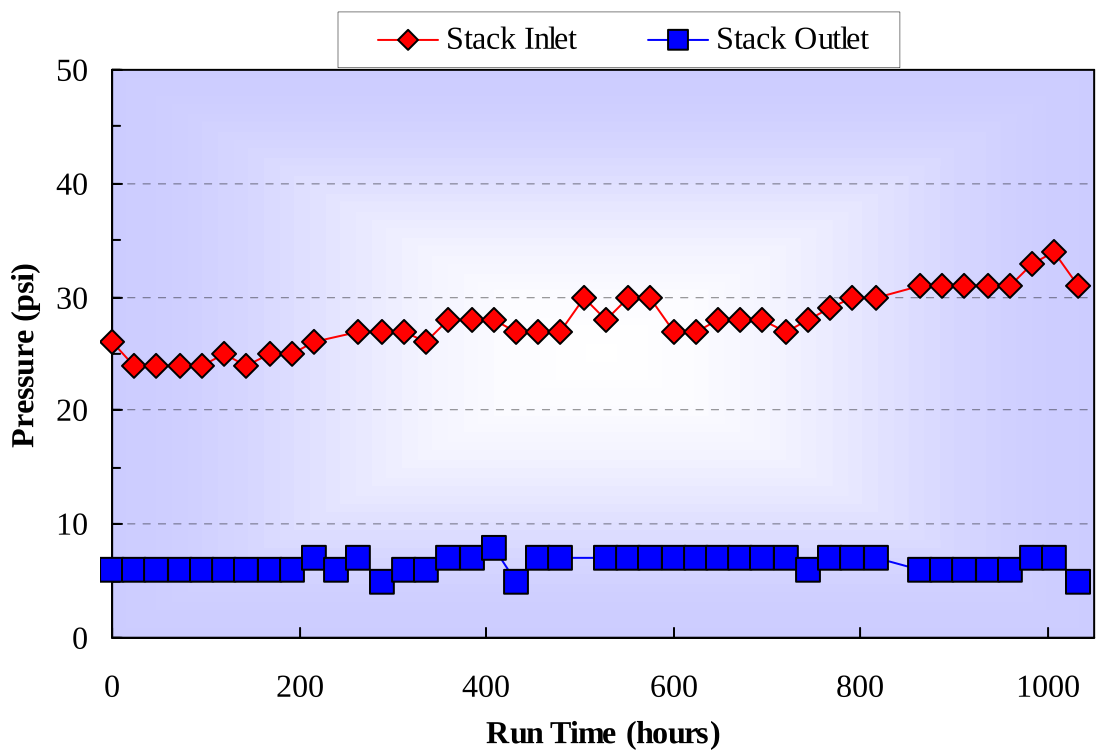

Figure 5-22. EDR pilot stack pressures – Phase I

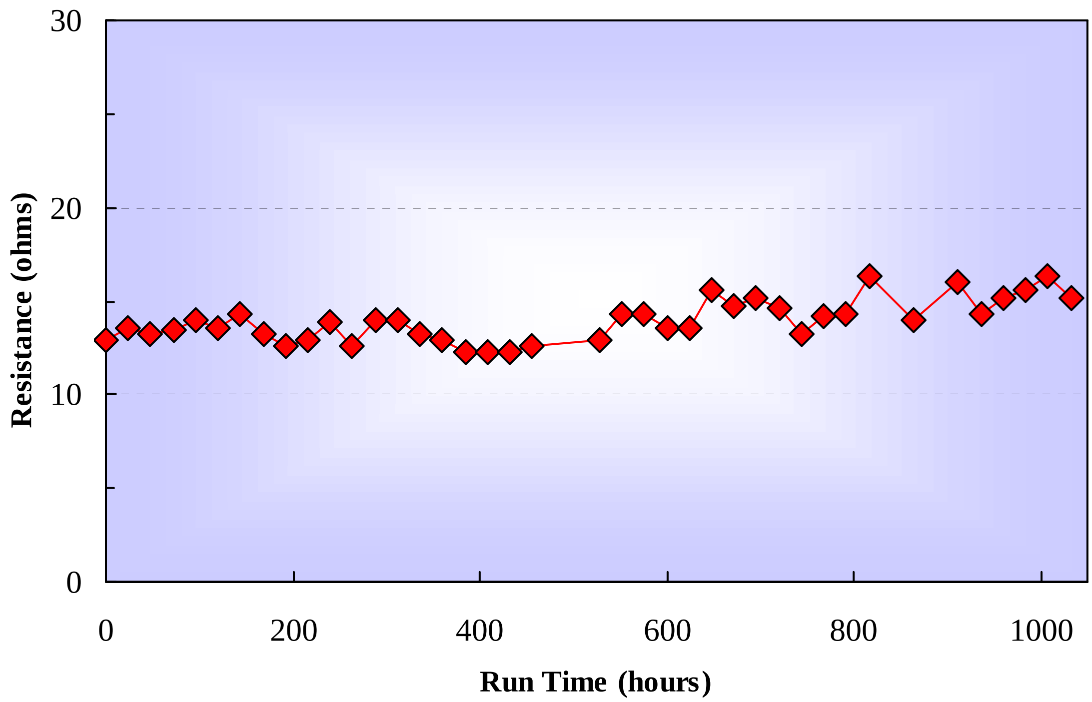

Figure 5-23. EDR pilot electrical resistance – Phase I

5.2.2 Phase II – Pretreatment Alternatives (500 mg/L TDS Treatment Goal)

A general process flow schematic for Phase II EDR pilot operation is provided in Figure 5-24. Tertiary treated wastewater from the SJ/SC WPCP was fed to the EDR pilot plant after being pretreated with either (a) GAC/MMF, (b) MF, or (c) cartridge filtration only. For the GAC/MMF pretreatment configuration, a small amount of the chlorinated feed water was allowed to enter the stacks to maintain a 0.5 to 1 mg/L free chlorine residual to prevent biological fouling on the EDR membranes. In the remaining configurations, no dechlorination of the feed water was performed. Final suspended solids removal was achieved using a 5-10 µm cartridge filter.

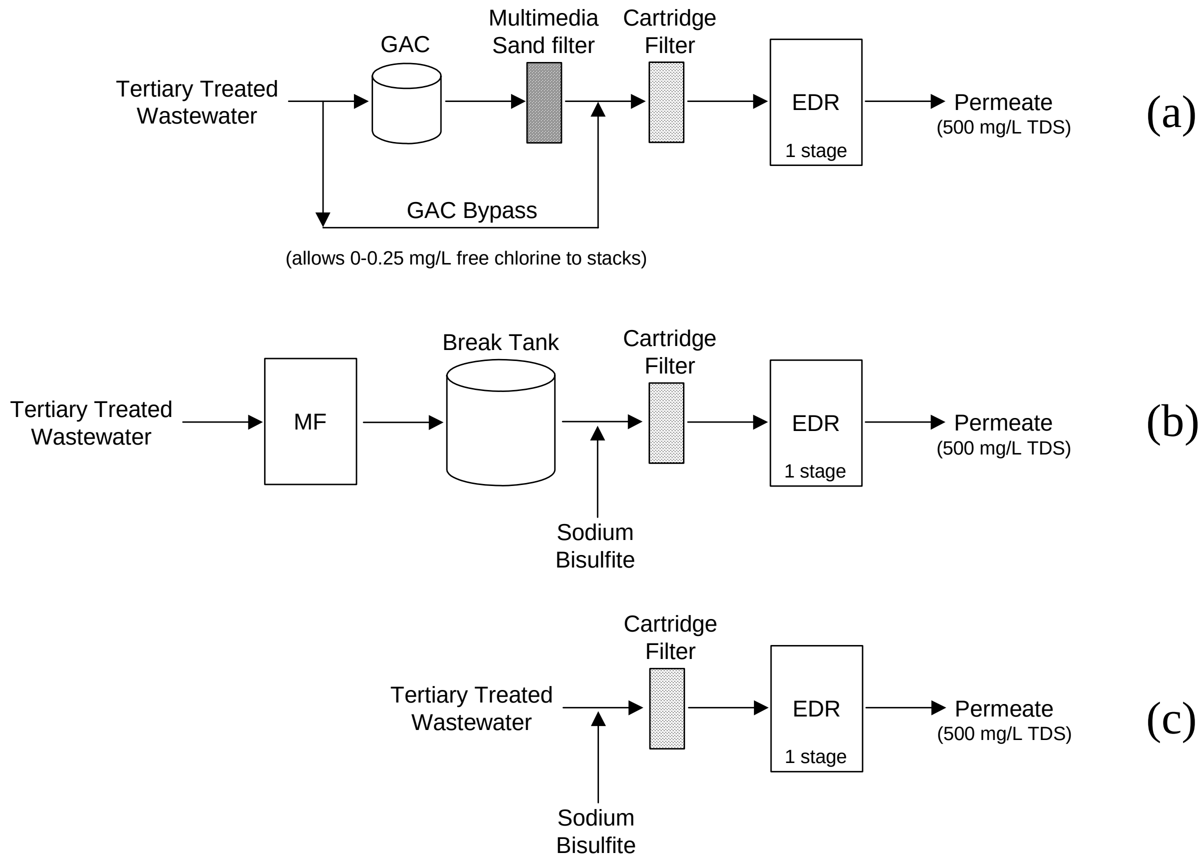

Figure 5-24. EDR pilot testing schematic (pretreatment alternatives)

In Phase II, the EDR pilot operated continuously for nearly 1300 hours, during which the three pretreatment configurations were evaluated. As shown in Figure 5-25, the EDR feed flow rate varied between 20 and 27, depending on which pretreatment option was being evaluated at the time. Figure 5-11 summarizes the operation and performance of the EDR in this test period. A lower flowrate, 20 gpm, was fed to the EDR during the MF pretreatment condition, because the MF permeate was divided between the EDR and RO. Table 5-2 summarizes specific water qualities and the amount of salt rejected for specific ions by the EDR.

Table 5-7. Summary of EDR operation and performance – Phase II

Parameter	GAC/MMF		MF (PVdF)		Cartridge Filtration	
	Range	Avg	Range	Avg	Range	Avg
Feed Water Flow Rate (gpm)	25 – 27.5	27	20 – 21	20	25 – 29	25
Product Flow Rate (gpm)	21 – 23	22	16 – 17	16.5	22 – 24	22
Feed Water Recovery (%)	78 – 92	82	75 – 86	81	86 – 89	88
Feed Water Temperature (ºC)	20 – 20	20	19.5 – 20	20	21 – 22	22
Feed TDS (mg/L)	675 – 790	750	707 – 780	740	670 – 770	760
Product TDS (mg/L)	500 – 590	550	475 – 550	550	450 –550	480
Stage 1 Voltage (V)	57 – 67	61	66 – 68	67	56 – 67	57
Stage 1 Current (amps)	3.3 – 4.1	3.8	3 – 4	3.5	2.6 – 3	2.8
Stack Inlet Pressure (psi)	32 – 38	34	24 – 28	24	43 – 50	50
Stack Inlet DP (inches H_2O)	78 – 98	82	60 – 84	61	10 – 26	12
Stack Outlet Pressure (psi)	5 – 7	5.4	5 – 7	6	7 – 9	7
Stack Outlet DP (inches H_2O)	32 – 80	60	50 – 60	60	10 – 17	10
Salt Rejection (%)	24 – 30	27	26 – 34	26	23 – 36	36

Table 5-8. Summary of EDR water quality – Phase II

Parameter	GAC/MMF		MF (PVdF)		Cartridge Filtration	
Parameter	**Feed (mg/L)**	**Product (mg/L)**	**Feed (mg/L)**	**Product (mg/L)**	**Feed (mg/L)**	**Product (mg/L)**
Chloride	188	130	195	130	195	121
Nitrate as N	8.6	6	9	5.7	8.4	5.4
Sulfate	98	57	100	55	100	63
Calcium	52	30	54	28	57	28
Magnesium	32	19	32	17	33	17
Silica	25	25	25	25	25	25
Sodium	140	120	150	120	158	123
Turbidity	1.0	0.8	0.1	0.1	1.0	0.6
TOC	8	7	7	6	9	8
Conductivity (µS)	1250	900	1200	830	1250	830
TDS (mg/L)	730	520	700	480	730	500

During all three pretreatment configurations, the EDR operated consistently at a feed water recovery of approximately 85 percent. The pilot unit utilized a single stack with one electrical stage to reduce the influent TDS to 500 mg/L, as shown in Figure 5-26. This consistent effluent water quality was maintained for entire Phase II test period and no operation failures occurred.

As previously discussed (Baseline Operation – Phase I), the pressure drop through the membrane stack was monitored to indicate if any membrane fouling was occurring. As shown in Figure 5-27, the stack outlet pressure was maintained below 10 psi for all conditions. During GAC/MMF pretreatment, the initial stack pressure represented a continuation of operation from Phase I EDR operation and a similar fouling rate was observed as the stack inlet pressure increased from 32 to 38 psi. The EDR operated for approximately 500 hours using GAC/MMF as a pretreatment step.

Next, MF pretreatment was used and the EDR was operated for an additional 500 hours during which the stack inlet pressure was observed to be lower, at approximately 25 psi. The lower stack inlet pressure was expected to be due to the lower feed flow rate applied to the EDR during this period of testing (20 gpm versus 27 gpm, previously). Additionally, the stable inlet pressure observed during operation with MF pretreated wastewater was most likely due to the superior water quality of the EDR feed, with respect to suspended solids. As shown in Table 5-8, the feed turbidity of the GAC/MMF pretreated water was about 1 NTU. Although very clean for tertiary effluent water quality, MF pretreated water was consistently measured below 0.1 NTU.

After the MF pretreatment test phase was complete, the EDR operated directly with reclaimed wastewater from the SJ/SC WPCP with only cartridge filtration. The purpose of this phase was to operate the EDR with a minimal amount of pretreatment (cartridge filtration only) to determine if the reclaimed water quality was sufficiently clean to avoid harmful fouling of the EDR membranes. Unfortunately, only 300 hours of continuous operation were achieved. An unexpected power surge in the electrical supply, seriously damaged the system, and prematurely terminated this evaluation. Based on the data collected, however, the stack inlet pressure began at 43 psi (most likely due to the higher flow rates used similar to the GAC/MMF operation) and approached the 50-psi limit by the end of operation. Typically, 50 psi is the recommended upper limit of operation, as recommended by the equipment manufacturer.

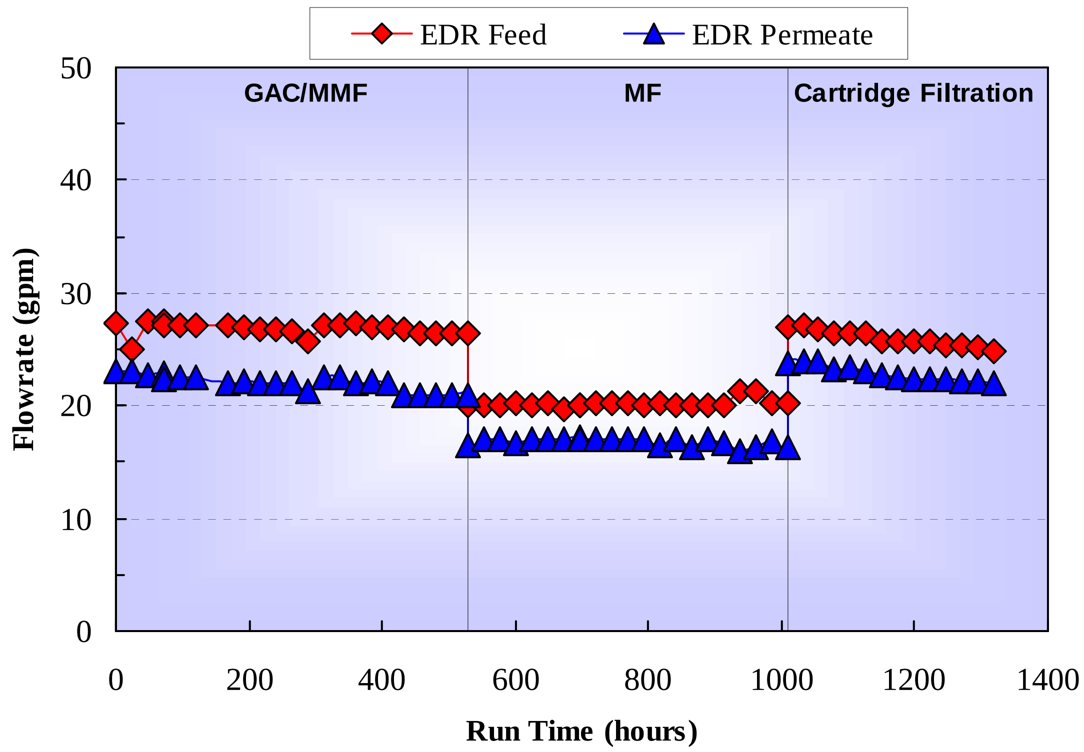

Figure 5-25. EDR pilot flowrates – Phase II

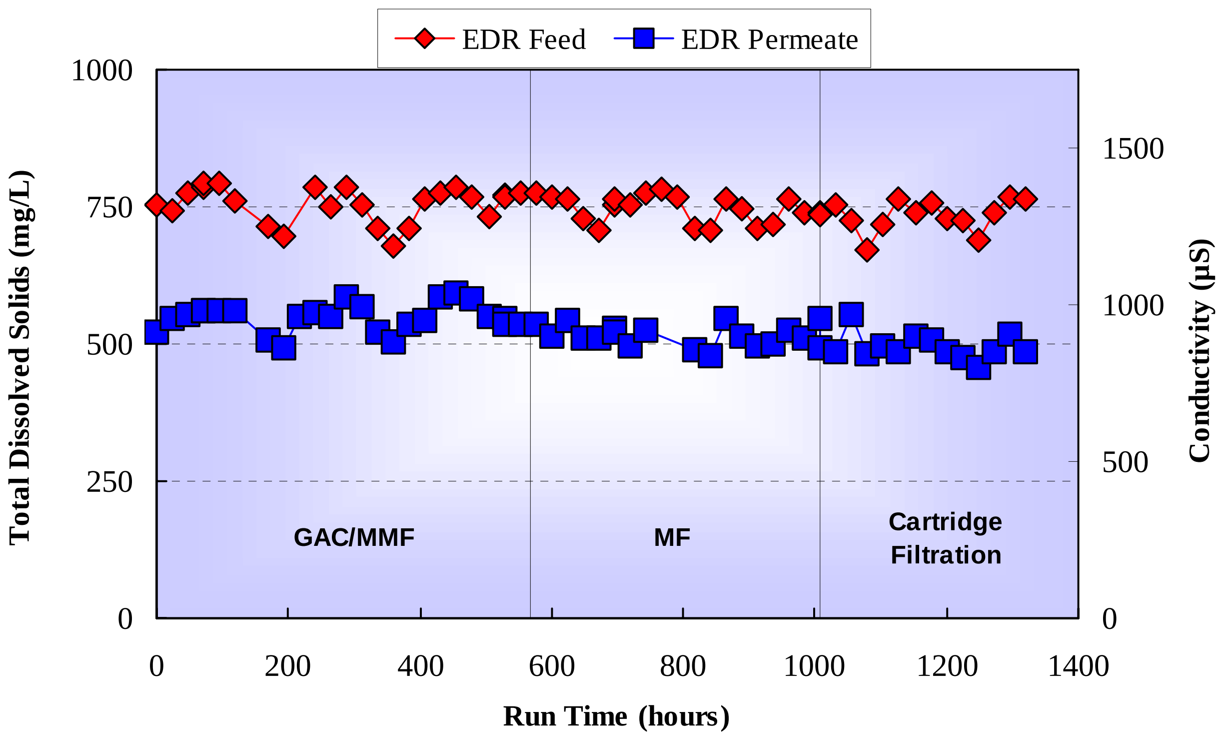

Figure 5-26. EDR pilot TDS levels – Phase II

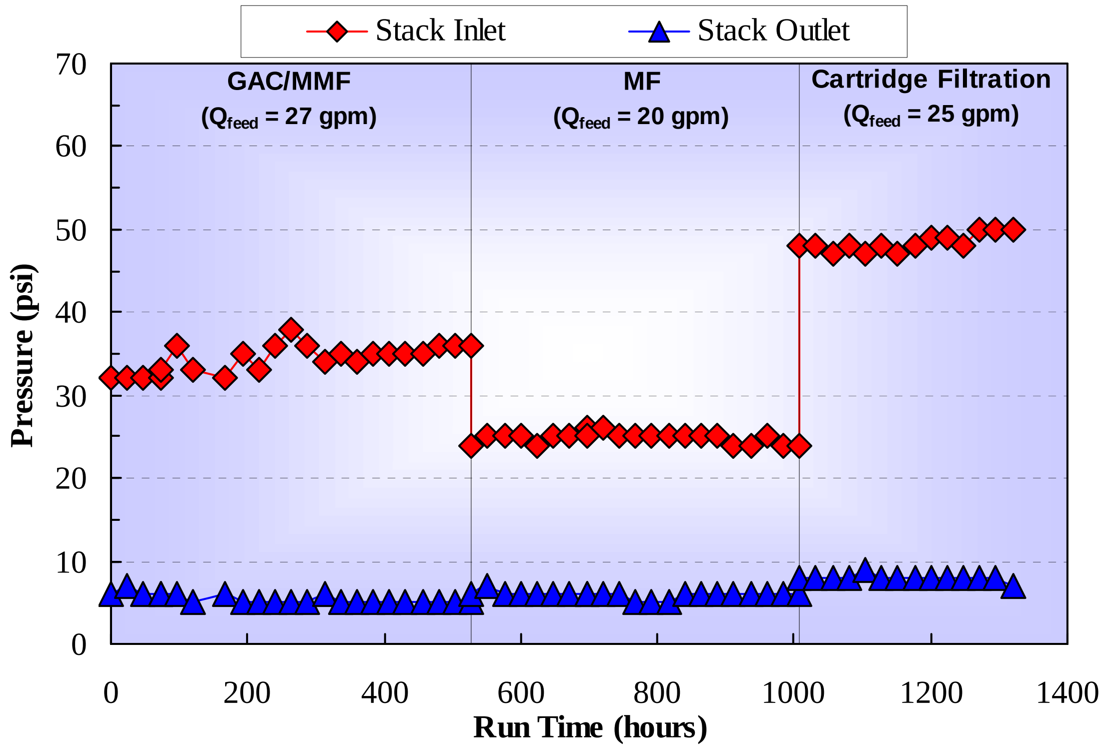

Figure 5-27. EDR pilot stack pressures – Phase II

In order to achieve TDS removal to 500 mg/L, the applied voltage was set to achieve the desired water quality goal. For the different flow rates, different voltages and resulting currents were required during each pretreatment condition to achieve the desired 500 mg/L effluent TDS level, shown Table 5-7. This variability makes it difficult to directly compare the three based on these parameters. The electrical resistance, however, provides a reasonable basis comparison for the three pretreatment conditions, in addition to providing insight about potential EDR membrane fouling.

Figure 5-28 shows that, for the entire test period (all pretreatment conditions), a constant increase in the electrical resistance occurred. Similar to the increase in the inlet stack pressures, a slight fouling trend was observed, however, neither pretreatment condition can be attributed to an accelerated fouling rate over the other. It has been assumed that the observed fouling rate was minimal and was characteristic of normal operation. During extended operation, recovery of performance would be achieved through regular maintenance and CIP cleans.

These important findings indicated that, when fed the high quality effluent from the SJ/SC WPCP, the EDR can be operated using a minimal amount of pretreatment without excessive fouling, operational failure, or a decrease in product water quality. Additionally, removing extraneous pretreatment equipment could provide a significant cost saving in operation as was discussed in Section 4.

It is important to note that despite the availability of high quality reclaimed water (i.e., low turbidity, low metals, etc.) during this pilot test, extra attention must be allowed to fully understand the impact of irregular and/or intermittent full-scale plant changes and their effect on the reclaimed water quality. Temporary changes made during full-scale operation of the SJ/SC WPCP to deal with unexpected events (i.e., high flow/demand, plant upsets, increased free chlorine, extra or alternative coagulant addition, etc.) could potentially create a harmful condition for the EDR membranes.

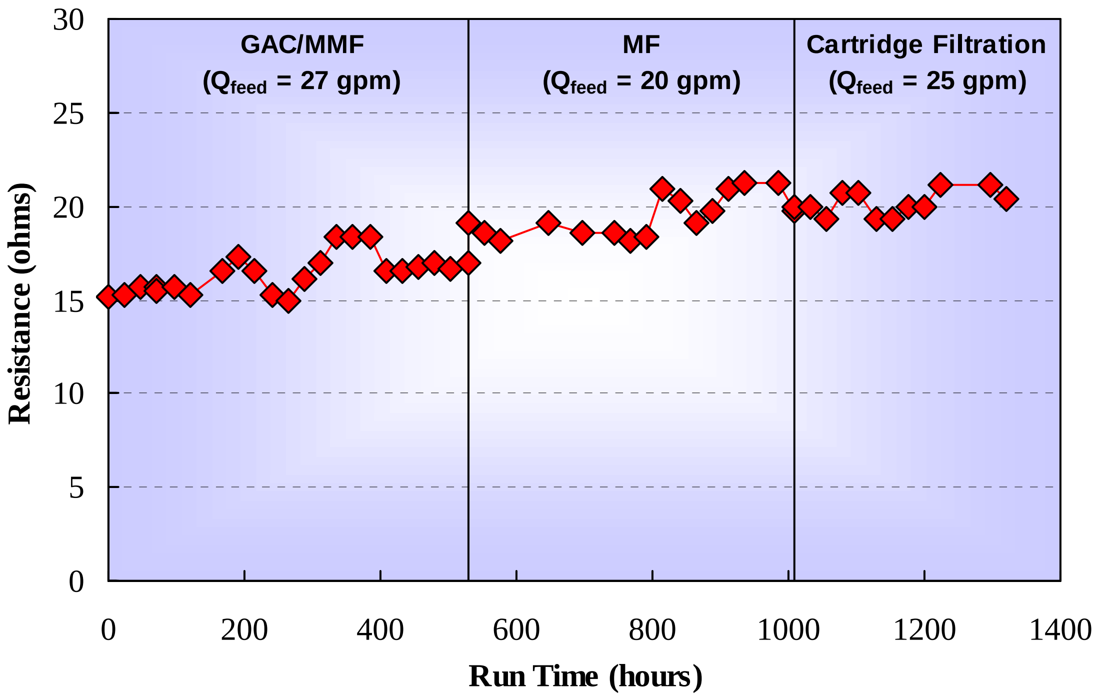

Figure 5-28. EDR pilot electrical resistance – Phase II

6. Process Comparison

The primary purpose of this study was to determine the feasibility of EDR and MF/RO advanced water treatment processesto reduce the salinity of the tertiary treated wastewater produced by the SJ/SC WPCP and to compare their cost-effectiveness. As noted in Section 5, both processes were found to be capable of reducing effluent salinity. This section compares the performance of the processes with respect to the following:

- **operational issues** including maintenance requirements and process failures experienced during pilot testing;
- **water quality** produced from each demineralization process; and
- **projected costs** for construction and operation of full-scale facilities required to produce 50 mgd of recycled water with a final TDS concentration of 350 and 500 mg/L.

6.1 Operation

The following operation issues were experienced during startup and operation of the pilot equipment. A detailed memo of operational experiences noted by plant operators of the SJ/SC WPCP is included in Appendix A.

Ionics EDR Operation

- <u>Algal fouling.</u> As noted in Section 4, initially sodium bisulfite was used to remove chlorine from the EDR system. Chemical addition was subsequently replaced by granular activated carbon (GAC) filters, and two GAC contactors were connected in series to remove chlorine from EDR feed water. The GAC contactors proved to be effective at removing free chlorine; in fact, there was evidence that complete dechlorination occurred in the first GAC contactor allowing algae to grow in the second GAC contactor, as indicated by inspection through the clear viewing ports. This eventually lead to biofouling of the EDR membranes, producing a dramatic increase in the inlet stack pressure. Corrective measures were taken by flushing the EDR stack with the undechlorinated EDR influent to remove algae, restoring the inlet operating pressure to 25 psi. To prevent further biofouling from occurring, a small stream of chlorinated feed water was allowed to bypass the GAC contactors to maintain a constant 0.5 to 1 mg/L free chlorine residual to the EDR stack.

USFilter MF/RO

- <u>CMF-S Membrane Damage.</u> During the initial phase of testing, polypropylene hollow-fiber membrane modules were chosen for the pretreatment of RO feed. These PP membranes are known to be sensitive to free chlorine concentrations and require dechlorination strategies to prevent damage. An ammonia feed system was installed to chloraminate the tertiary treated wastewater feed water and protect the PP membranes. Midway through the MF/RO pilot testing, the ammonia feed system failed causing the PP membrane fibers to break and become damaged. As a result, the CMF-S was not able to effectively reduce the SDI below 5 causing RO membrane fouling and reduced salt rejection efficiencies. Although the first phase of MF/RO pilot testing was prematurely terminated, sufficient data was collected, as presented in this report to characterize MF/RO operation under baseline conditions. Upon further consideration, the Project Team and USFilter decided 1) to replace the polypropylene membranes with polyvinylidene fluoride membranes more tolerant of free chlorine, and (2) to clean and replace RO membrane elements as necessary and test their integrity to ensure continued performance.

6.2 Water Quality

Pilot testing results demonstrated the EDR and MF/RO treatment could reduce the TDS of recycled water to 350 mg/L and 10 mg/L, respectively. Based on the pilot EDR configuration, the treatment goal could be achieved without additional blending of the product water with untreated recycled water. The RO product water, however, would need to be blended to achieve the appropriate treatment goal. It is often common for full-scale demineralization plants (both EDR and MF/RO) to blend some source water with the permeate of the plant. An advantage is gained by blending because plant sizes are often reduced, resulting in lower capital and operating costs (AWWA 1999).

Table 6-1 compares the water quality of EDR permeate to MF/RO permeate blended with source water. For this example, the MF/RO permeate was normalized to the average TDS of the EDR permeate. It can be seen in this table that similar water quality is achieved for both processes. Slightly higher concentrations of sulfate, calcium, and magnesium would be present in a MF/RO blended product quality. Silica and sodium, however, were still maintained below EDR product water. It is important to note that this water quality comparison was based on pilot-scale testing results.

Table 6-1. EDR versus estimated blended MF/RO pilot product water quality

Parameter	Unit	Feed	RO Blended Product [1]	EDR Product
Cl^-	mg/L	188	87.0	88
NO_3-N	mg/L	7.1	3.3	3.4
SO_4	mg/L	96	44.7	33
Br^-	mg/L	<1.0	<1.0	<1.0
NO_2-N	mg/L	<0.05	<0.05	<0.05
Al	mg/L	0.06	0.1	<0.005
Ba	mg/L	0.020	0.01	0.007
B	mg/L	0.510	0.3	0.481
Ca	mg/L	59.1	27.2	10.8
Cr (Total)	mg/L	<0.002	<0.002	<0.002
Fe	mg/L	0.07	0.1	0.05
Mg	mg/L	31.7	14.6	7.11
SiO_2	mg/L	24.0	11.2	23.5
Na	mg/L	156	73.3	101
Sr	mg/L	0.387	0.2	0.078
NH_3-N	mg/L	<0.1	<0.1	<0.1
Conductivity	umhos/cm	1250	591	614
pH	SU	7.3	6.5	6.9
TOC	mg/L	9	5.2	6
TKN	mg/L	0.4	<0.3	<0.3
TSS	mg/L	<2	<2	<2
Turbidity	NTU	0.7	0.4	0.5
Hardness, total (CaCO3)	mg/L	250	115	55
Alkalinity, total (CaCO3)	mg/L	190	90	120
TDS	mg/L	750	350	350
UV-254	Abs/cm	0.109	0	0.068

[1] - Based on pilot operating RO flux=15 gfd, FWR=65%

6.3 Treatment Costs

Pilot testing of both EDR and MF/RO demonstrated that both processes were effective at demineralizing tertiary treated wastewater produced by the San Jose/Santa Clara SJ/SC WPCP. As a result, a cost analysis was performed to estimate capital and operational costs associated with a full-scale EDR and MF/RO system with a production capacity of 50 mgd. Both Ionics and USFilter were contacted to provide costs associated with their respective systems. Design criteria used to estimate advanced treatment costs were based on information collected during pilot testing and manufacturer recommendations. The cost estimate encompassed the following criteria:

- Treatment capacity of 50 mgd
- Construction related and labor costs
- Product water (including blending water) to achieve effluent with 350 and 500 mg/L TDS
- Operation and maintenance (O&M) costs including consumables, power and parts based on pilot performance
- Membrane system capital costs (obtained from Ionics and USFilter).
- Ancillary (pre)treatment costs

Construction related costs include engineering design, site work, legal, and administrative work involved in the construction of a new brine treatment facility. These values are calculated using a range of construction related costs (as a percentage of the capital cost), based on experience with water treatment plant construction (Table 6-2).

Table 6-2. Range of construction related costs

Construction Related Cost	Average Range*
Civil Site Work	1 to 10%
Instrumentation	3 to 15%
Electrical Site Work	7 to 12%
Piping	5 to 12 %
Construction Contingency	10 to 35%
Contractor Overhead and Profit, Bonds, and Insurance	10 to 20%
Engineering, Legal and Administrative	10 to 30%

* Average ranges based on experience with surface water treatment plant design.

Construction contingencies are applied to the cost estimate to account for items not specifically included in a project scope but found to be necessary. The level of contingency selected should reflect the level of detail provided during pre-design. A low contingency budget reflects a high degree of confidence in the pre-design and a high contingency budget reflects a low level of confidence in the pre-design. A low degree of confidence may be due to limited availability of detailed costs or an experimental treatment technology. The recommended contingency levels for the varying types of cost estimates are listed in Table 6-3. A 20 percent contingency was applied to the engineering analysis in this study.

Table 6-3. Recommended contingency for corresponding level of estimate

Type of Cost Estimate	Level of Accuracy	Recommended Contingency
Order-of-Magnitude	+50% to –30%	20% to 30%
Conceptual	+40% to –20%	20% to 15%
Preliminary Design	+30% to –15%	15% to 10%
Definitive	+15% to –5%	10% to 5%

6.3.1 Conceptual Level Full-Scale Treatment Costs

A preliminary conceptual cost analysis was performed using capital and O&M costs provided by equipment manufacturers and a MWH cost model of MF/RO treatment facilities (DRIP 2004). These full-scale EDR and MF/RO treatment costs were developed for a 50-mgd system including blending of product water with untreated tertiary effluent to achieve a finished water TDS of both 350 and 500 mg/L.

Based on results of pilot-scale testing, three treatment scenarios were evaluated:

- RO using MF pretreated water
- EDR using MF pretreated water
- EDR without pretreatment (cartridge filtration only)

Design Criteria

As GAC for chlorine reduction is considerably more expensive than chemical (i.e., SBS reduction) a separate cost for GAC pretreatment was not considered. The following assumptions Table 6-4) were used in the cost analysis of MF/RO full-scale plants designed to produced 50-mgd of recycled water. Additionally, the blending ratios used to achieve 350 and 500 mg/L TDS recycled water are outlined in Figure 6-1.

Table 6-4. MF/RO design conditions used for conceptual cost analysis

Parameter	350 mg/L TDS Goal	500 mg/L TDS Goal
RO Flux	12 gfd	12 gfd
RO FWR	85 percent	85 percent
MF Flux	20 gfd	20 gfd
MF FWR	90 percent	90 percent
MF Feed Flow	43 mgd	21 mgd
RO Permeate Flow	35 mgd	17 mgd
Blending Flow Rate	15 mgd	33 mgd
RO Product TDS	20 mg/L	20 mg/L
Blending Water TDS	750 mg/L	750 mg/L

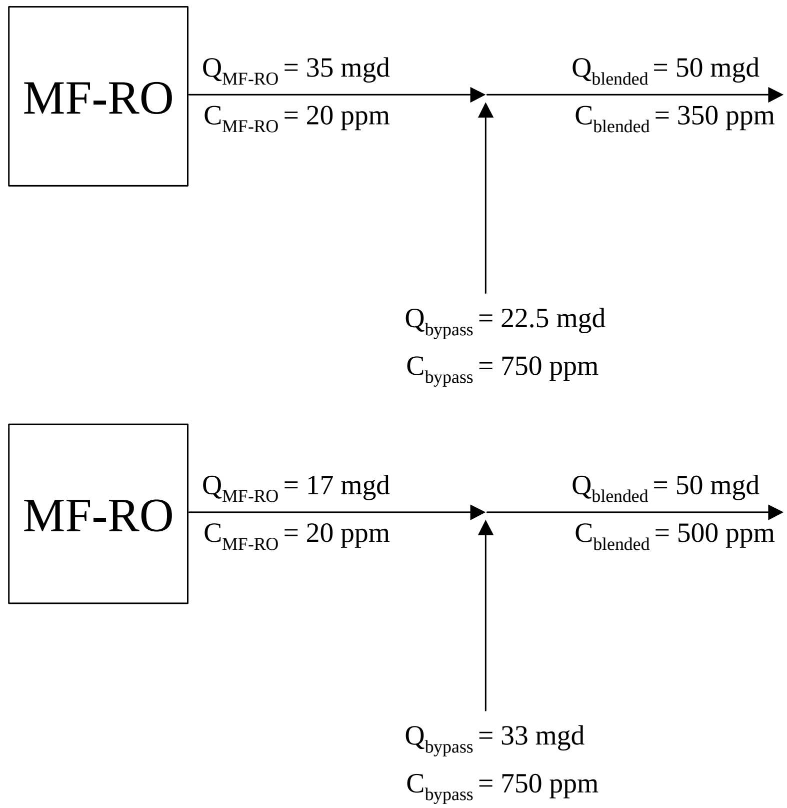

Figure 6-1. Schematic of MF/RO 50-MGD blending at 350 and 500 mg/L treatment goals

The full-scale EDR facility was based on using a 3-stage design with a total feed water recovery of 85 percent. Based on pilot-scale testing results, it was determined that the high quality reclaimed water produced by SJ/SC WPCP might be suitable for operation of an EDR system without any pretreatment. It is important to note that additional pilot testing of the EDR without pretreatment is recommended to confirm the long-term performance of the EDR membranes, since operation during this test program was focused on evaluating multiple pretreatment options (with short term test periods) to determine if any immediate fatal flaws would be encountered. Additionally, EDR testing at the “Cartridge Filtration” condition was prematurely terminated due to a pilot plant power failure. The following assumptions (Table 6-5) were used in the cost analysis of MF/RO full-scale plants designed to produced 50-mgd of recycled water. Additionally, the blending ratios used to achieve 350 and 500 mg/L TDS recycled water are outlined in Table 6-2.

In order to provide perspective regarding the estimated cost of full-scale EDR treatment, additional costs were developed to include MF pretreatment. A cost estimate of MF/EDR would provide the high range for the treatment costs as compared to the EDR without pretreatment. Similar to the MF/RO cost estimate, the EDR costs were estimated for the production of 50 mgd blended product for both 350 and 500 mg/L TDS treatment goals by utilizing appropriately sized systems at 34 and 19 mgd production rates, respectively (Figure 6-2).

Table 6-5 EDR design criteria used for conceptual cost analysis

Parameter	350 mg/L TDS Goal	500 mg/L TDS Goal
EDR operation	3-stage	3-stage
EDR FWR	85 percent	85 percent
EDR Feed Flow	40 mgd	22.3 mgd
EDR ProductFlow	34 mgd	19 mgd
Blending Flow Rate	19.6 mgd	31.2 mgd
RO Product TDS	82 mg/L	82 mg/L
Blending Water TDS	750 mg/L	750 mg/L

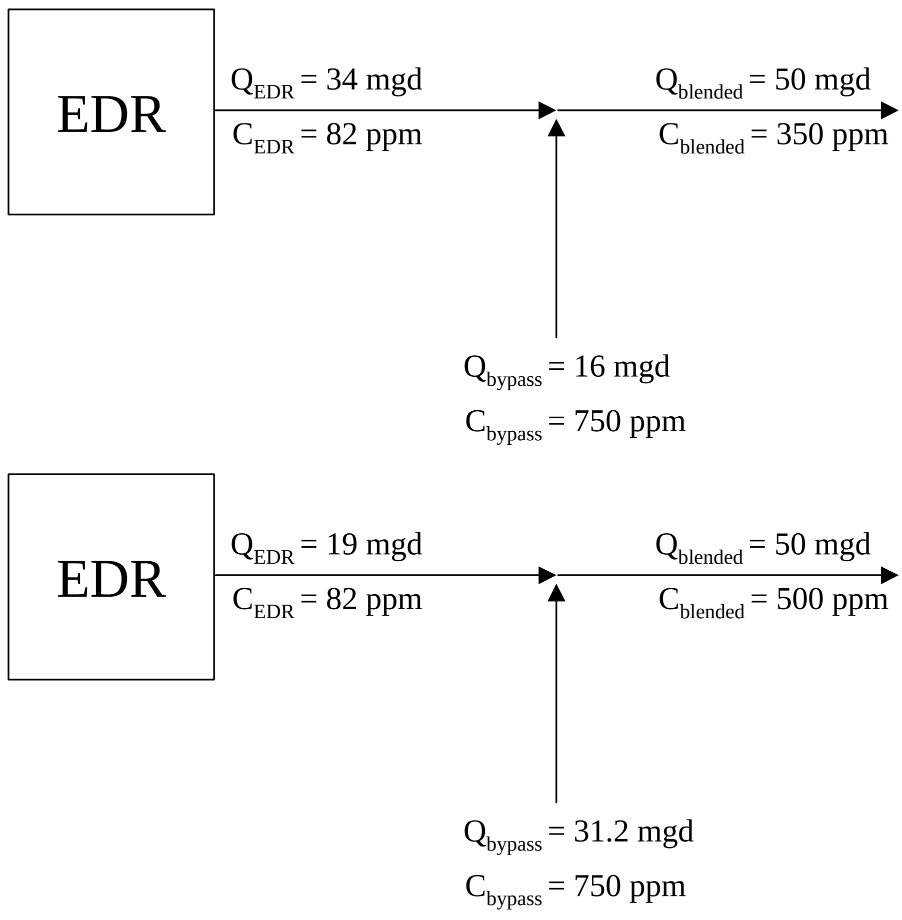

Figure 6-2. Schematic of EDR 50-MGD blending at 350 and 500 mg/L treatment goals

Estimated Treatment Costs

As shown in Table 6-7, the conceptual cost of treating 750 mg/L TDS water to produce 50 mgd of 350 mg/L TDS was $0.57/kgal and $0.86/kgal for EDR and MF/RO, respectively. A significant cost savings was estimated for the EDR without required pretreatment. Once MF pretreatment was factored into the EDR costing, the cost of treatment was $0.85/kgal and was similar to the cost of MF/RO ($0.86/kgal).

At the 500 mg/L TDS treatment goal, a significant cost savings for all treatment scenarios would be expected because of the reduction in the size of the treatment plant required. Treatment costs for MF/RO and MF/EDR systems were estimated to be $0.51/kgal and 0.55/kgal, respectively. This represents approximately a 40 percent reduction of the overall treatment costs. Similarly, a full-scale EDR plant without pretreatment is estimated to only cost $0.32/kgal. A breakdown of the estimated capital and O&M costs is presented in Table 6-6.

Table 6-6. MF/RO versus EDR full-scale cost breakdown

MF/RO Percent Breakdown		**EDR Percent Breakdown**	
Capital		*Capital*	
MF Cost	21 %		
RO Cost	26 %	EDR Cost	60 %
Building	17 %	Building	15 %
Site Work	23 %	Site Work	20 %
Miscellaneous	13 %	Miscellaneous	5 %
O&M		*O&M*	
RO Power	13 %	EDR Power	39 %
Other Power	9 %		
Chemicals	6 %	Chemicals	2 %
Membrane Replacement	21 %	Membrane Replacement	18 %
Labor/Maintenance	35 %	Labor/Maintenance	25 %
Cartridge Filter	1 %	Cartridge Filter	8 %
Miscellaneous	15 %	Miscellaneous	8 %

Table 6-7. Conceptual life cycle costs for desalting treatment options

		Total Capital Cost ($ million)	Annual O&M Cost ($ million)	Total Annualized Treatment Cost[1,2] ($/1000 gallons)
350 mg/L TDS				
MF/RO	35 mgd	$102	$6.8	$0.86
MF/EDR	34 mgd	$73.6	$9.0	$0.85
EDR	34 mgd	$33.4	$7.4	$0.57
500 mg/L TDS				
MF/RO	17 mgd	$62.2	$3.8	$0.51
MF/EDR	19 mgd	$52.9	$5.5	$0.55
EDR	19 mgd	$20.2	$4.2	$0.32

[1] Amortization for 20 years at 5 percent interest

[2] Annualized cost normalized to 50 mgd blended product with 350 mg/L TDS

6.4 Process Advantages

Based on the conceptual cost comparison of the two pilot tested demineralization processes, it was shown that EDR may have a significant cost advantage over MF/RO, if no pretreatment beyond cartridge filtration is required. Factoring in a MF pretreatment step to the EDR, causes both advanced treatment technologies to be cost-competitive. Additional process advantages of both EDR and MF/RO treatment processes may be realized in light of overall treatment costs. The following Table 6-8 summarizes the advantages and disadvantages of both technologies, as compared to one another.

Table 6-8. Advantages and disadvantages of demineralization processes

EDR	RO
Advantages ❑ Minimal pretreatment may be required (cartridge filtration is recommended) ❑ Operates at a low pressure ❑ Process is much quieter because high pressure pumps are not required ❑ Antiscalant is not required ❑ Membrane life expectancy is longer because foulants continuously are removed during the reversal process ❑ Requires less maintenance than RO due to reversal process *Disadvantages* ❑ Limited to 50 percent salt rejection for a single membrane stack (stage) ❑ Requires larger footprint to produce similar quantity and quality of water if multiple staging is used ❑ Electrical safety requirements ❑ Less experience for wastewater demineralization in the U.S. ❑ Not as effective at removing microorganisms and many anthropogenic organic contaminants	*Advantages* ❑ RO membranes provide a barrier to microorganisms and many anthropogenic organic contaminants (for the treated portion of the water produced) ❑ More demonstrated experience for wastewater demineralization ❑ RO membranes can remove more than 90 percent of TDS ❑ Source water blending will reduce size of systems ❑ Flexibility to provide higher quality water, if desired. *Disadvantages* ❑ Requires high pressure to achieve high salt rejection ❑ Requires pretreatment processes to minimize scaling and fouling. ❑ Requires chemical addition for MF & RO fouling control ❑ More routine maintenance may be required to maintain performance

REFERENCE LIST

American Water Works Association. Reverse Osmosis and Nanofiltration. Manual of Water Supply Practices – M46 , First Edition. 1999

Chou, Lynn. Telephone Interview. City of San Diego. January 2004.

Desalination Research and Innovation Partnership (DRIP). "An analysis of the cost-effectiveness of desalting recycled water by non-thermal technologies." September 2002.

Desalination Research and Innovation Partnership (DRIP). "Cost of salt removal from various waters with membrane systems." June 2004.

MWH. Water Repurification Project: Membrane Prequalification Pilot Study. Prepared for the City of San Diego. 1997.

Reahl, Eugene. Telephone Interview. Ionics, Inc. August 2004.

Wilf, M., and Schierach,M.,K., Improved performance and cost reduction of RO seawater systems using UF pretreatment, Desalination, 135, (2001) 61-68.